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Process Design Manual for Upgrading Existing Wastewater Treatment Plants

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NOTICE

The mention of trade names or commercial products in this publication is for illustration purposes and does not constitute endorsement or recommendation for use by the U. S. Environmental Protection Agency.

ABSTRACT

The main purposes of this manual are to examine situations that necessitate upgrading of existing municipal wastewater treatment plants and to discuss and evaluate the corrective actions that are required to upgrade these existing plants. Upgrading to overcome organic and hydraulic overloadings and/or to meet more stringent treatment requirements is considered.

The manual emphasizes that operational improvement and modifications to existing unit operations be considered as the logical initial approach to upgrading existing treatment plants, before major expansion of existing facilities is implemented.

Because of the numerous alternatives available for upgrading an existing treatment plant, it is necessary to understand thoroughly the fundamentals of the various unit operations commonly used in municipal wastewater treatment plants. Therefore, this manual examines in depth the capabilities, limitations, and interrelationships of the various unit processes. The manual also examines hypothetical situations requiring upgrading of unit operations and describes "order of magnitude" costs associated with the upgrading of various unit operations.

One chapter of the manual presents case histories of upgrading of existing wastewater treatment plants to illustrate the approaches actually used in these circumstances.

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FOREWORD

The formation of the United States Environmental Protection Agency marked a new era of environmental awareness in America. This Agency's goals are national in scope and encompass broad responsibility in the area of air and water pollution, solid wastes, pesticides, and radiation. A vital part of EPA's national water pollution control effort is the constant development and dissemination of new technology for wastewater treatment.

It is now clear that only the most effective design and operation of wastewater treatment facilities, using the latest available techniques, will be adequate to meet the future water quality objectives and to ensure continued protection of the nation's waters. It is essential that this new technology be incorporated into the contemporary design of waste treatment facilities to achieve maximum benefit of our pollution control expenditures.

The purpose of this manual is to provide the engineering community and related industry a new source of information to be used in the planning, design and operation of present and future wastewater treatment facilities. It is recognized that there are a number of design manuals, manuals of standard practice, and design guidelines currently available in the field that adequately describe and interpret current engineering practices as related to traditional plant design. It is the intent of this manual to supplement this existing body of knowledge by describing new treatment methods, and by discussing the application of new techniques for more effectively removing a broad spectrum of contaminants from wastewater.

Much of the information presented is based on the evaluation and operation of pilot, demonstration and full-scale plants. The design criteria thus generated represent typical values. These values should be used as a guide and should be tempered with sound engineering judgment based on a complete analysis of the specific application.

This manual is one of several available through the Technology Transfer Office of EPA to describe recent technological advances and new information. This particular manual was initially issued in October of 1971 and this edition represents the first revision to the basic text. Future editions will be issued as warranted by advancing state of the art to include new data as it becomes available, and to revise design criteria as additional full-scale operational information is generated.

CHAPTER 1

INTRODUCTION

The ability of wastewater treatment plants to perform at required levels of efficiency becomes more critical as water pollution abatement programs achieve their objectives. Deviations from design performance, which were formerly of lesser consequence, now become paramount because of their impact on the receiving waters. Improved process monitoring and plant operation will obviously reduce the incidence of inadequate performance, but many cases are the result of more basic deficiencies in the treatment system. Upgrading of wastewater treatment plants may be required to handle higher hydraulic and organic loadings to meet existing effluent quality and/or to meet higher treatment requirements. Any of these situations requires optimization of existing facilities before consideration of additional treatment facilities. It is necessary that a distinction be made between upgrading to accommodate higher hydraulic and organic loads, and upgrading to meet stricter treatment requirements. Existing facilities can be made to handle higher hydraulic and organic loads by process modifications, whereas meeting higher treatment requirements usually requires significant expansion and/or modification of existing facilities.

Regardless of the cause, the result is that an inadequately treated effluent is discharged. The historical solution to such a problem has been plant expansion along the same lines as the original facility, or addition of conventional unit processes to add secondary or tertiary treatment to the system. Depending on its application, a generalized approach such as this does not necessarily make optimum use of the previously existing facilities nor of the expanded facilities. The situation is further complicated where regional treatment systems are proposed for the future and existing facilities are inadequate for the interim period. In such cases, a solution must make optimum use of available technology, with minimum capital expenditure.

Rapid urbanization, development of industries, and stricter treatment requirements often necessitate unanticipated upgrading of treatment plants or premature implementation of upgrading programs. Many existing treatment plants are not capable of meeting the more stringent performance levels required by today's water quality standards. In addition, there are needs for interim improvements. These considerations, plus economic pressures to optimize pollution abatement expenditures, make it mandatory that a logical and technically sound approach to upgrading existing treatment facilities be established. This is especially true because of the numerous alternatives available for consideration prior to the selection of a method for upgrading an existing facility. It is for this reason that a major plant expansion, i.e., complete duplication of existing unit treatment processes, for the purposes of this manual will be considered the least attractive upgrading procedure available, since this approach does not consider optimization of existing facilities.

Therefore, the purpose of this manual will be to present necessary information for considering various courses of action with regard to an impending or existing plant overload situation, or with regard to increasing the efficiency to meet stricter treatment standards. The diversity of causes that necessitate upgrading of existing plants precludes the use of this manual as a conventional design manual. Therefore, it is aimed at establishing a framework of possible alternative methods of upgrading overloaded treatment plants. To facilitate the information presented, only plants treating "typical" domestic wastewaters will be considered.

Particular upgrading procedures are stressed as interim methods which may be implemented with a minimum amount of effort and capital expenditure prior to a more elaborate upgrading or even a major plant expansion. Cost information has been compiled and estimates prepared for the upgrading of individual unit processes. Capital costs are based on an EPA-cost index of 175, and unless noted otherwise do not include an allowance for land, right-of-ways, engineering and legal fees, contingencies or interest during construction. When available, cost information has also been included for the reported case histories on plant upgrading. Due to the varying complexity of existing plants, the real benefit of the subsequent cost information will be as a tool for developing comparative capital costs for various upgrading techniques. Particular unit process cost information must be used cautiously, since the complexity of the individual situation will dictate the costs required for upgrading.

The aspects of nutrient removal, although extremely important and oftentimes responsible for upgrading action at many treatment plants, will be discussed only briefly herein since more detailed information is presented in other manuals either currently available or soon to be published.

CHAPTER 2

INVESTIGATIVE APPROACH

2.1 Purpose of Upgrading Existing Wastewater Treatment Plants

Upgrading of existing wastewater treatment plants may be required for a number of reasons, including the following:

1. To meet more stringent treatment requirements
2. To increase hydraulic and/or organic loading capacity
3. To improve poor performance due to improper plant design and/or operation

The approach to the solution of an upgrading problem will vary depending upon the objectives to be attained. In some cases, upgrading only a portion of the plant may achieve significant improvements in plant performance. This chapter describes the general investigative techniques applicable to most upgrading situations.

2.2 Identification of Existing Problem Areas

As the initial step in solving an upgrading problem, the engineer should thoroughly familiarize himself with the following aspects of the existing facility:

1. Efficiency of treatment
2. Normal operational and maintenance procedures
3. Condition of structures
4. Condition of process hardware
5. Staffing pattern and level of operator skill.

Thorough discussions with plant operating personnel are of great benefit in determining the above. These discussions must be supplemented by a complete review of plant operating records such as daily flow charts, operating logs and laboratory data. Normally, much of this information is compiled on standard State regulatory agency forms which are submitted monthly.

The engineer should become completely familiar with the type of sampling and flow measurement techniques employed by the plant personnel and verify their accuracy. One difficult area in assessing plant performance data is the reliability of influent flow measurement data. Often, plant flow data are obtained from some type of flow recording instrument, and reliable flow information is possible only when the treatment plant operator routinely calibrates the flow measuring and recording instruments. A representative portion of the operating data, normally a period of one to three years, should be evaluated by the engineer. If adequate operational and performance data are not available, it is the responsibility of the engineer to collect the additional information needed to complete the evaluation.

2.3 Upgrading to Meet More Stringent Treatment Requirements

Increased pressure from Federal and State governments, and a more ecologically minded public, are requiring local communities and sanitary districts to enforce existing water quality standards. In addition, many regulatory agencies are stipulating (1) increased BOD and SS removal, (2) the maintenance of minimum DO concentrations in the plant effluent or receiving body of water, (3) more stringent disinfection requirements, (4) phosphorus removal, (5) partial or complete oxidation of ammonia nitrogen, and, in some cases, (6) high levels of nitrogen removal.

An essential feature of any upgrading plan is the provision of adequate flexibility in plant design to accommodate future treatment requirements. These requirements will vary depending on the location of the treatment plant, the expected useful lifetime of the existing facility, and the condition and projected use of the receiving body of water. Regardless of the ultimate treatment goal, adequate consideration must be given to maximum utilization of the existing plant in the overall upgrading plan. In some instances, moderate improvements in plant efficiency may be sufficient, while in other situations, extensive improvements may be required to meet very stringent effluent requirements. Often, upgrading to meet long-term water quality goals involves expansion of major portions of the plant to accommodate increased flows.

Since this type of upgrading involves a major capital investment, a detailed evaluation of all feasible alternatives is justified. This evaluation should include a careful analysis of the increased operating cost and staffing pattern required by the upgraded facility along with the municipality's ability to provide the needed financial support. Often, major plant expansion and added tertiary facilities will substantially increase the cost of sludge handling and disposal, which represents a significant portion of the total plant operating cost.

2.4 Upgrading to Relieve Hydraulic and Organic Overloads

The performance of different unit processes within a treatment plant is affected to varying degrees by hydraulic and organic overloads. The relationship between higher flows and the corresponding changes in BOD and SS loading is also an important consideration. For example, a significant increase in flow without a corresponding increase in organics will require a different upgrading approach than when the increase in flow is accompanied by an increase in BOD and SS.

Hydraulic and organic overloading normally occurs as the tributary population increases beyond the plant design capacity. Projections must then be made for a reasonable design period into the future. Forecasting changes in population and wastewater flows in connection with upgrading of an existing treatment plant may be quite burdensome but generally will not be subject to as much uncertainty as in similar forecasting for a relatively undeveloped area. In many cases, the maximum anticipated growth is defined by saturation of the tributary area. Potential extension of this area must also be considered and is often limited by topographical constraints and political boundaries. Analysis of local area growth patterns; examination of local influences such as land use planning studies, zoning regulations and wastewater discharge ordinances; and full use of State, County and local planning agencies can all be extremely useful in judging the future expected flows for a given upgrading situation.

Rapid industrialization in an existing plant's service area can cause increased hydraulic and organic loads. The alert community, before issuing a building permit, should notify industries that pretreatment may be required for wastewaters containing toxic or treatment inhibiting materials, or for wastewaters having an unusually high percentage of organic or inorganic material compared to typical domestic wastewaters. Equalization of industrial wastewater discharges may be helpful in minimizing diurnal flow variations and in damping short term, high strength discharges.

Population equivalent equates the organic content or flow contribution of industries to the ordinary per capita contribution present in domestic wastewaters. Even a relatively small industry may contribute a significantly higher organic loading than the existing domestic load. Population equivalents for many industrial wastewaters should be based on COD analysis rather than BOD, since extremely strong or toxic wastes may show an artificially low BOD.

Historical data on raw wastewater flows and concentrations can provide valuable information on the magnitude, frequency and duration of peak flows and loadings. Electronic data processing and plotting techniques are effective in reducing data to a usable form and can provide a firm base for projecting future loading conditions (1). Such analyses can detect whether significant seasonal or weekly variations exist. This is particularly important in determining whether plant expansions can be designed for annual average values, or whether workday industrial loads warrant higher than annual average design values for sizing of plant components.

2.4.1 Hydraulic Overloads

Excessive hydraulic loadings in primary and secondary clarifiers are a major cause of reduced plant efficiency. Increased surface overflow rates and reduced detention periods in primary clarifiers result in poor BOD and SS capture which increases the loads to be handled by subsequent treatment units. SS removal in secondary clarifiers is even more sensitive to excessive hydraulic loadings than primary clarifiers. High secondary clarifier surface overflow rates in activated sludge plants can result in the direct carry-over of biological solids from the clarifier sludge blanket into the effluent. In trickling filter plants, high secondary clarifier loadings prevent effective capture of the biological solids sloughed from the filter media.

In any upgrading situation, it is essential that a thorough study be made of plant influent flow records. It is desirable to review actual meter flow charts to determine typical diurnal flow variations. These data can be used to size flow equalization basins, as discussed in Chapter 3. Moreover, such analyses are required to determine whether seasonal or continuous excessive flows exist, and whether the excessive flows are due to groundwater infiltration or to inflow contributions.

Since excessive flows can hydraulically overload the unit processes in a wastewater treatment plant, every effort should be made to reduce excessive infiltration/inflow contributions before undertaking upgrading techniques. Excessive infiltration/inflow is a likely possibility when influent BOD and SS concentrations are consistently below 150 mg/l. Plant flow and connected population data, and in some cases records of potable water usage, can be used to determine if an infiltration/inflow problem exists. Major causes of infiltration are leaky manholes, faulty lateral connections and leaky pipe joints. Major sources of excessive inflow are illegal downspouts, footer drains, cross-connections with storm sewers and surface runoff into the top of illogically placed manholes.

Where it has been established that an infiltration/inflow problem exists, detailed study is recommended (2). Usually this consists of an initial analysis to determine the magnitude and location of the problem area. A survey is then made to inspect and/or test the suspected locations. Usually, the affected portions of the sewers are cleaned before the inspection is begun. Typical testing and inspection techniques include:

1. Smoke testing

2. Hydrostatic testing

3. Air testing

4. Closed circuit television

5. Photographic methods

The survey report should identify the specific locations for remedial action, the recommended rehabilitation method and an estimate of the rehabilitation cost. This cost can then be compared to the cost of upgrading the plant with and without infiltration/inflow control to establish a cost effective course of action. The most common remedial techniques considered are:

1. Internal or external pressure grouting with chemical sealants

2. Manhole grouting

3. Replacing, elevating and/or sealing of manhole covers

4. Replacement of severely damaged sewer sections or service connections

5. Insertion of sewer liners

6. Removal or plugging of illegal inflow connections such as downspouts or footer drains.

Reported costs for inspection and repair of sewers vary widely due to a number of factors such as:

1. Sewer age, construction materials and construction methods

2. Accessibility of sewers

3. Sewer size

4. Soil condition

Because of the great variety of conditions which can be encountered, average cost values or specific cost data can be misleading. However, a general cost of $5 to $30 per foot has been estimated (3).

A number of successful infiltration studies and control programs have been reported (3)(4)(5), and many others are ongoing. Once the problems are corrected, a continuing program of flow analysis and sewer inspection can significantly prolong the effective lifetime of the upgraded plant.

3.4.3 Organic Overloads

The more common causes of organic overloading of existing plants are the addition of substantial industrial loads, increased discharge of septage to the plant, and in-plant recycle from poorly operating or overloaded sludge processing operations. Recycled BOD and SS loads from properly loaded and operated thickeners, digesters, elutriation tanks and vacuum filters can amount to as much as 25 percent of the total plant organic load. When these solids processing units become overloaded or are improperly operated, recycled BOD and SS loads can reach 50 percent or more of the total plant organic load. Unless pretreated or equalized, or unless the plant was originally sized to account for them, these intermittent recycle loads can significantly decrease overall plant performance and, when extremely severe, can cause complete process failure.

Overloads created by industrial contributors can be relieved by pretreatment, by equalization or by expansion of critical plant components. As discussed earlier, Federal and State regulations, along with local ordinances, may necessitate pretreatment. Equalization of industrial flows by the discharging industry, or at the municipal plant, is especially effective when the discharge is of relatively short duration.

Excessive BOD loadings usually do not have a major adverse effect on primary treatment. However, in biological treatment systems, BOD overloads can cause anaerobic conditions in aeration tanks and trickling filters, and may result in unstable, unsettleable activated sludges which are prone to bulking and other problems. High influent BOD loads will also result in excessive secondary sludge production. The engineer evaluating an existing plant should check all of the normally used loading parameters to determine if additional tank volume, oxygen supply, and return and waste sludge capacity is necessary, or if operating adjustments can be implemented.

Excessive SS loadings will result in increased primary and secondary sludge production which, in turn, may exceed the capacity of existing thickeners, digesters and dewatering equipment. If adequate plant data exist, the actual loading parameters can be compared to the plant design values to determine the degree of overloading and to identify alternative courses of action.

Complete system mass balances for phosphorus and SS are excellent tools to determine the validity of wastewater and sludge flow measurements and analytical data. Solids balances are a key factor in identifying the source and magnitude of excessive in-plant recycling loads. Phosphorus balances are useful because phosphorus is the one major pollutant common to all municipal wastewaters which does not break down (even partially) to gaseous forms during biological treatment.

In smaller plants located in communities served partially by septic tank systems, the discharge of relatively large quantities of septage can create significant BOD and SS overloads and sludge handling and odor problems. Thus, in such an upgrading evaluation, the quantity of septage which will be discharged to the plant should be thoroughly investigated. In northern states, where tanks are normally emptied during the spring and summer months, the annual quantities of materials collected will be introduced to the plant over a 6- to 8-month period. The quantity and characteristics of septage vary widely (5)(6)(7)(8)(9). This variation is greater in communities that do not employ a permit system, or otherwise regulate the collection and disposal of septage. In these cases, septage haulers will indiscriminately include septic tank contents along with raw wastewater collected from pit toilets, wastes from camping trailer pump-out stations, waste motor oil from service stations, cutting oil and other hard to treat or toxic wastes from small industries throughout the community.

Typical concentration ranges for septage are shown in Table 2-1 (7)(8).

TABLE 2-1
CHARACTERISTICS OF SEPTAGE

Parameter	Range	Average
BOD, mg/l	2,000 - 25,000	9,000
TSS, mg/l	7,000 - 106,000	40,000
VSS, percent of TSS	47 - 82	69
COD, mg/l	5,000 - 80,000	54,000

Two methods are commonly employed to handle septage at existing treatment facilities. The method of choice will depend on a complete analytical characterization of the septage, the relative volume of septage to influent wastewater flow and physical constraints at the plant site.

If the ratio of septage to total plant flow is relatively small on a pound of BOD/day basis, these wastes may be discharged to an aerated storage basin at the plant site and metered into the plant at a controlled rate so as not to seriously interfere with normal plant operation and performance.

If the ratio of septage to total plant flow is high, or if the septage contains toxic or treatment inhibiting substances, separate treatment facilities must be provided. Full-scale separate treatment techniques that have been considered include aerobic digestion followed by sludge dewatering and aerobic digestion followed by introduction of digester contents at a controlled rate into the plant. Separate treatment pilot studies have included aerobic digestion, followed by dewatering on sand beds, polymer thickening, and high lime (pH 11.5) stabilization followed by dewatering on sand beds (7).

2.5 Upgrading to Improve Plant Design and Operation

Many existing plants, although well designed, have not produced the anticipated performance because of inadequate operator staffing and training, and/or a lack of financial support and commitment to water pollution abatement from the community. On the other hand, some poorly designed plants are functioning well because of the excellence of plant operation. Each upgrading situation will contain its own innate characteristics relative to plant design and operation. The design engineer must carefully evaluate the interrelationships of the design and its operation to arrive at the solution which optimizes both of these factors.

2.5.1 Plant Design

In the past, the problems associated with inadequately designed municipal wastewater treatment plants have been a major concern of Federal and State agencies. For this reason, most states have adopted conservative design guidelines and review procedures which must be followed unless the engineer can substantiate a less conservative viewpoint. The implementation of these procedures by regulatory agencies has reduced the frequency of inadequate plant design. However, there are many aspects of detailed design other than those covered by these regulations which can still adversely affect overall plant performance. Usually, the plant operators responsible for day-to-day supervision of plant systems and components can identify major design deficiencies.

Among the numerous design features which can adversely affect plant operation and performance are:

1. Inadequate standby equipment
2. Fixed speed units where variable speed is essential, such as raw wastewater pumps, recycle pumps and air blowers
3. Poor hydraulic and solids distribution among identical units operating in parallel
4. Insufficient or inflexible return and sludge wasting pumping capacity
5. Inability of instrumentation and equipment to operate at low flows and loads occurring in the early lifetime of the plant

6. Lack of tank dewatering systems to permit rapid servicing of submerged equipment

7. Lack of flexibility in disinfection systems to permit prechlorination for odor control, or return sludge chlorination for control of sludge bulking

8. Inadequate electrical and hydraulic capacity to permit standby pumps and air blowers to operate in parallel

9. Inadequate laboratory facilities

10. Lack of accessibility to key mechanical equipment for routine maintenance or repair

From a process standpoint, the most common deficiency is plant design based on average flow and BOD and SS loadings, with no recognition of peak conditions. In many plants, sustained peak influent flows and BOD and SS loads can reach two or more times average values. Therefore, all plant units and systems, including primary and secondary sludge pumping, air supply equipment and solids processing components, should be designed to perform successfully at peak conditions. Frequently, peak flows and BOD and SS mass loadings do not occur simultaneously (10). Thus, a design based on the concurrence of peak flow and peak pollutant concentration conditions may result in excess capacity. Analysis of existing plant data, if available, will assist in determining the appropriate peak mass loadings and flows.

2.3.2 Plant Operation

One of the primary considerations in evaluating an overloaded plant is in the area of plant operation and control. Therefore, no physical upgrading should be considered before the engineer is assured that the plant is being operated to yield its maximum efficiency.

A major tool required for proper process control is frequent and accurate sampling and laboratory analysis. The locations at which samples are taken are critical, particularly in solids-laden flows where solids tend to travel near the bottom of the conduit.

Many wastewater treatment plants do not have a laboratory equipped to analyze wastewater samples from the various treatment units in order to assess their performance. Improper operation, coupled with inadequate laboratory control, increases the probability of poor treatment. For this reason, plants must be staffed with an adequate number of competent operators and laboratory personnel. Further, sufficient funds must be made available for proper maintenance and for the purchase of adequate sampling and analytical equipment and supplies.

Because of the availability of field training programs (11) and the trend to mandatory operator certification, operator skills have improved markedly over the past five years. The engineer involved in an upgrading situation should identify any areas in which additional training would result in an improved operation.

2.6 Consideration of Applicable Upgrading Techniques

Technology in the field of wastewater treatment in the past decade has provided many innovative upgrading procedures. Various research projects sponsored by the U. S. EPA have resulted in a better understanding of existing and new unit processes. In addition, new types of equipment have significantly enlarged the range of alternatives available for plant upgrading.

It has long been recognized that the performance of a wastewater treatment plant is affected by variations in the influent flow. Equalization of extreme flows can dampen the fluctuations in flows and mass loadings to existing plants.

Various techniques such as chemical coagulation, polyelectrolyte addition and novel settling devices are being successfully used to increase removal efficiency in primary and secondary clarifiers. These procedures can improve removal efficiency while maximizing hydraulic throughput in the existing facilities.

Several modifications of the conventional activated sludge process, including step aeration, contact stabilization and complete mix have been successfully used to upgrade various treatment plants. A most significant recent development in the activated sludge treatment process is the use of oxygenation as a substitute for air aeration.

Another method of upgrading an overloaded secondary plant is to provide additional treatment ahead of the existing biological treatment facilities. The use of synthetic media trickling filters should be considered when roughing treatment would reduce BOD loadings on existing facilities to acceptable operating levels. Synthetic media filters have been used successfully as roughing filters in industrial wastewater treatment, and it is very likely that they will be used in the future for upgrading municipal treatment plants.

The beneficial effect of a nitrified effluent on DO in receiving waters is receiving increasing attention. For this reason, some regulatory agencies are requiring nitrification of treatment plant effluents during summer periods, and in some cases are contemplating a year-round nitrification requirement. Nitrification at higher wastewater temperatures may be accomplished through modifications to the existing treatment units, such as increasing aeration tank volume or adding chemicals to the primary clarifier to decrease the organic loading to existing aeration units. However, dependable year-round nitrification will normally require a two-stage biological treatment system.

In an increasing number of instances, treatment plants which are functioning satisfactorily (design flow not exceeded) will have to improve BOD and SS removals because of more stringent water quality standards. The necessary additional treatment can often be achieved by polishing the treatment plant effluent. Several methods are currently available and have been used successfully, including tertiary aerobic and facultative lagoons, microscreening, multimedia filtration and activated carbon treatment.

Various sludge handling developments which have been successful are: (1) high-rate anaerobic digestion, (2) aerobic digestion, (3) air flotation thickening or centrifugal thickening of waste activated sludge, (4) polyelectrolyte addition to improve thickening and dewatering of sludges and (5) heat treatment processes for sludge stabilization.

Although this manual presents in detail current technology available for upgrading existing wastewater treatment plants, the engineer should keep in mind the following overall considerations which will affect the economics of upgrading:

1. The physical condition of existing plant equipment and structures and their potential uses in an upgrading situation.
2. The length of time before a major expansion will be required, based on population and wastewater flow projections.
3. The time required for implementation of various upgrading techniques.
4. The compatibility of upgrading procedures with future planned expansion.
5. The financial assistance available to the community.
6. The costs of the various upgrading techniques that can be used to achieve essentially the same result. The operation and maintenance costs, as well as the capital costs, may vary substantially for different available alternatives. Therefore, economic comparison of these alternatives is essential.

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CHAPTER 3

FLOW EQUALIZATION

3.1 Introduction and Concept

3.1.1 General

The cyclic nature of wastewater flows in terms of volume and strength is well recognized. Nearly all municipal wastewater treatment plants today are processing variable wastewater flows. However, improved efficiency, reliability and control is possible when physical, biological and chemical processes are operated at or near uniform conditions. For this reason, flow equalization is employed in the field of water supply and in the treatment of some industrial wastewater. Presently, the advent of more demanding water quality standards is stirring interest in the application of flow equalization to municipal wastewater treatment.

The primary objective of flow equalization basins for municipal treatment plants is simply to dampen the diurnal flow variation, and thus achieve a constant or nearly constant flow rate through the downstream treatment processes. A desirable secondary objective is to dampen the concentration and mass flow of wastewater constituents by blending the wastewater in the equalization basin. This results in a more uniform loading of organics, nutrients and other suspended and dissolved constituents to subsequent processes.

Through achieving these objectives, flow equalization can significantly improve the performance of an existing treatment facility, and is a useful upgrading technique. In the case of new plant design, flow equalization can reduce the required size of downstream facilities.

3.1.2 Variations of Flow Equalization

Equalization of municipal wastewater flows may be divided into three broad categories:

1. Equalization of dry weather flows
2. Equalization of wet weather flows from separate sanitary sewers
3. Equalization of combined storm and sanitary wastewater.

This chapter is primarily concerned with equalization of dry weather flows. This procedure provides a technique for achieving normal operation of a treatment plant under near ideal loading conditions. Its relatively low cost makes it attractive for upgrading an overloaded plant.

Increased wet weather flows in sanitary sewers is the sum of two components, infiltration and inflow. In some cases, it is feasible to equalize stormwater inflow, depending on its magnitude and duration. Infiltration from high groundwater tables can seldom be equalized. Equalization of wet weather flows from combined storm and sanitary sewers usually requires very large storage basins. The design of equalization basins to deal with these types of flow requires a special knowledge of the collection system, precipitation patterns, topography, and other factors not directly related to wastewater treatment. Strictly speaking, wet weather and combined sewer flow equalization cannot be considered as a wastewater treatment upgrading technique, and the design of such a facility is beyond the scope of this chapter. However, the concepts presented for dry weather flow equalization are generally applicable to equalization of wet weather and combined wastewater flows.

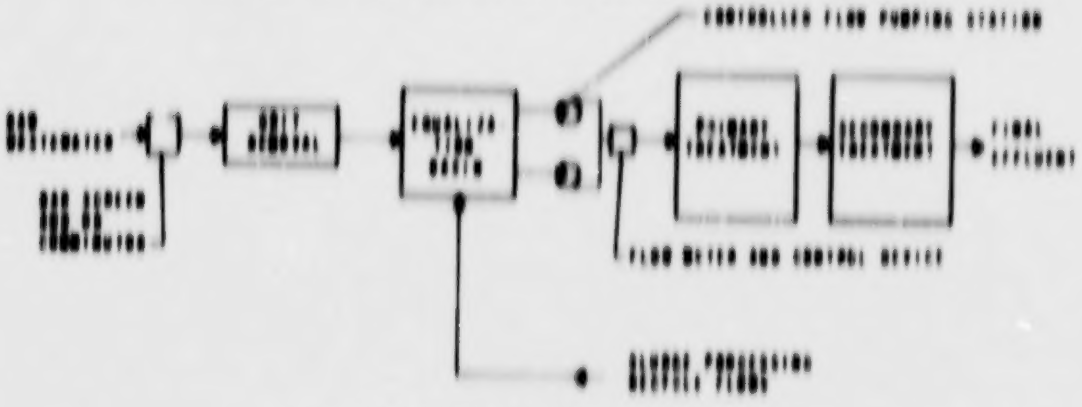
In some instances, large interceptor sewers entering the treatment plant can be effectively used as storage basins to dampen peak diurnal dry weather flow variations. In such cases, nightly or weekly drawdown of the interceptor system is necessary to flush out solids which may have been deposited during the previous storage period.

Although the use of influent sewers for equalization should not be ignored, the most positive and effective means to maximize the benefits possible with equalization is through the use of specially designed equalization basins. These basins should normally be located near the head end of the treatment works, preferably downstream of pretreatment facilities such as bar screens, comminutors, and grit chambers. Adequate aeration and mixing must be provided to keep the basins aerobic and prevent solids deposition.

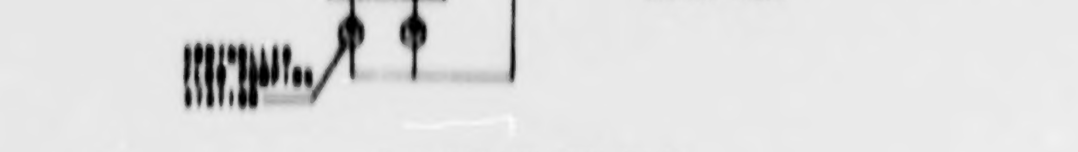
It is sometimes desirable to locate the equalization basin at strategic locations within the collection system. This offers the added advantage of economically relieving trunk sewer overload during peak flow periods (1). However, it does result in the need for a pumping facility and therefore is best located where a need for pumping already exists.

Equalization basins may be designed as either in-line or side-line units. In the in-line design shown on Figure 2-1a, all the flow passes through the equalization basin. This results in significant concentration and mass flow damping. In the side-line design shown on Figure 2-1b, only that amount of flow above the daily average is diverted through the equalization basin. This scheme minimizes pumping requirements at the expense of less effective concentration damping.

FIGURE 3.1
SCHEMATIC FLOW DIAGRAMS OF EQUALIZATION FACILITIES


CONTROLLED FLOW PUMPING STATION
RAW WASTEWATER
BAR SCREEN AND/OR COMMINUTOR
GRIT REMOVAL
EQUALIZATION BASIN
PRIMARY TREATMENT
SECONDARY TREATMENT
FINAL EFFLUENT
FLOW METER AND CONTROL DEVICE

3-1a IN-LINE EQUALIZATION


OVERFLOW STRUCTURE
RAW WASTEWATER
BAR SCREEN AND/OR COMMINUTOR
GRIT REMOVAL
FLOW METER AND CONTROL DEVICE
EQUALIZATION BASIN
PRIMARY TREATMENT
SECONDARY TREATMENT
FINAL EFFLUENT
CONTROLLED FLOW PUMPING STATION

3-1b SIDE-LINE EQUALIZATION

For new construction and for upgrading large plants, it is desirable to construct compartmentalized or multiple basins. This feature will allow the flexibility to dewater a portion of the facility for maintenance or equipment repair while still providing some flow equalization. Where a basin is designed for storage and equalization of wet weather flows, compartmentalized tanks will allow the utilization of a portion of the basin for dry weather flow equalization.

Single basin installations may be used for upgrading small plants, but must have the provision to be dewatered while maintaining complete treatment. This will require a bypass line around the basin to allow the downstream portion of the plant to operate unequalized when the flow equalization facility is out of service.

3.2 Benefits of Dry Weather Flow Equalization

Flow equalization has a positive impact on all treatment processes from primary treatment to advanced waste treatment.

3.2.1 Impact on Primary Settling

The most beneficial impact on primary settling is the reduction of peak overflow rates resulting in improved performance, and a more uniform primary effluent quality. Flow equalization permits the sizing of new clarifiers based on equalized flow rates rather than peak rates. In an existing primary clarifier that is hydraulically overloaded during periods of peak diurnal flow, equalization can reduce the maximum overflow rate to an acceptable level. A constant influent feed rate also avoids hydraulic disruptions in the clarifier created by sudden flow changes, especially those caused by additional wastewater lift pumps suddenly coming on line.

LaGrega and Keenan (2) investigated the effect of flow equalization at the 1.8 mgd Newark, New York Wastewater Treatment Plant. An existing aeration tank was temporarily converted to an equalization basin. They compared the performance of primary settling under marginal operating conditions, with and without equalization. The results are shown in Table 3-1.

It has been demonstrated (3) (4) that preaeration can significantly improve primary settling, as discussed in Chapter 8. Roe (4) concluded that preaeration preflocculates SS thereby improving their settling characteristics. Indications are that this benefit may be realized by aerated equalization basins. This benefit may be diminished when the equalized flow is centrifugally pumped to the primary clarifier, due to the shearing of the floc.

TABLE 3-1
EFFECT OF FLOW EQUALIZATION ON PRIMARY SETTLING
NEWARK, NEW YORK

	Normal Flow	Equalized Flow
Primary Influent SS, mg/l	136.7	128.0
Primary Effluent SS, mg/l	105.4	68.0
SS Removal in Primaries, percent	23	47

Note: Average flow slightly higher in unequalized portion of study.

3.2.2 Impact on Biological Treatment

As contrasted to primary treatment or other mainly physical processes where concentration damping is of minor benefit, biological treatment performance can benefit significantly from both concentration damping and flow smoothing. Concentration damping can protect biological processes from upset or failure from shock loadings of toxic or treatment inhibiting substances. Therefore, in-line equalization basins are preferred to side-line basins for biological treatment applications.

Improvement in effluent quality due to stabilized mass loading of BOD on biological systems treating normal domestic wastes has not been adequately demonstrated to date. It is expected that the effect will be significant where diurnal fluctuations in organic mass loadings are extreme. This situation may arise at a wastewater treatment plant receiving a high-strength industrial flow of short duration. Damping of flow and mass loading will also improve aeration tank performance where aeration equipment is marginal or inadequate in satisfying peak diurnal loading oxygen demands (8).

The optimum pH for bacterial growth lies between 6.5 and 7.5. In-line flow equalization can provide an effective means for maintaining a stabilized pH within this range.

Flow smoothing can be expected to improve final settling even more so than primary settling. In the activated sludge process, flow equalization has the added benefit of stabilizing the solids loading on the final clarifier. This has two ramifications:

1. The MLSS concentration can be increased thereby decreasing the F/M and increasing the SRT. This may result in an increased level of nitrification, and a decrease in biological sludge production. It may also improve the performance of a system operating at an excessively high daily peak F/M.

2. Diurnal fluctuations in the sludge blanket level will be reduced. This reduces the potential for solids being drawn over the weir by the higher velocities in the zone of the effluent weirs.

3.2.3 Miscellaneous Benefits

In chemical coagulation and precipitation systems using iron or aluminum salts, the quantity of chemical coagulant required is proportional to the mass of material to be precipitated. Damping of mass loadings with in-line equalization will improve chemical feed control and process reliability, and may reduce instrumentation complexity and costs.

Flow smoothing will reduce the surface area required and enhance the performance of tertiary filters. A constant feed rate will lead to more uniform solids loadings and filtration cycles.

The equalization basin provides an excellent point of return for recycled concentrated waste streams such as digester supernatant, sludge dewatering filtrate and polishing filter backwash.

Some BOD reduction is likely to occur in an aerated equalization basin. A 10 to 20 percent reduction has been suggested (6) for an in-line basin equalizing raw wastewater. However, the degree of reduction will depend upon the detention time in the basin, the aeration provided, wastewater temperature and other factors. For an existing treatment plant, a simple series of oxygen uptake studies on a representative sample of wastewater can determine the BOD reduction that will occur.

Roe (3) observed that preaeration may improve the treatability of raw wastewater by creating a positive oxidation-reduction potential, thereby reducing the degree of oxidation required in subsequent stages of treatment.

3.3 Determination of Equalization Requirements

The design of an equalization basin requires the evaluation and selection of a number of features as follows:

1. In-line versus side-line basins
2. Basin volume
3. Degree of compartmentalization
4. Type of construction - earthen, concrete or steel

5. Aeration and mixing equipment

6. Pumping and control concept

7. Location in treatment system.

The design decisions must be based on the nature and extent of the treatment processes used, the benefits desired, and local site conditions and constraints.

It may not be necessary to equalize the entire influent flow where high flow or concentration variations can be attributed to one source, such as an industry. In these cases the desired benefits can be achieved by simply equalizing the industrial flow. This can be accomplished through construction of an equalization basin at the industrial site or through in-house industrial process modifications to effect an equalized wastewater discharge.

3.3.1 Determination of Required Volume

Two methods are available for computing equalization volume requirements. One procedure is based on the characteristic diurnal flow pattern. In this case, the function of the basin is to store flows in excess of the average daily flow and to discharge them at times when the flow is less than the average. The required volume can be determined graphically through the construction of a hydrograph. The second procedure is based upon the mass loading pattern of a particular constituent. This method computes the volume required to dampen mass loading variations to within a preset acceptable range (7) (8).

Since the prime objective of flow equalization in wastewater treatment is to equalize flow, the determination of equalization volume should be based on the hydrograph. Once the volume has been determined for flow smoothing, the effect on concentration and mass load damping can be estimated. The required volumes for side-line and in-line basins will be identical. The hydrograph procedure is discussed below.

The first step in design involves the establishment of a diurnal flow pattern. Whenever possible, this should be based upon actual plant data. It is important to note that the diurnal pattern will vary from day to day, especially from weekday to weekend and also from month to month. The pattern selected must yield a large enough basin design to effectively equalize any reasonable dry weather diurnal flow. Figure 3-2 depicts a typical diurnal flow pattern. The average flow rate is 4.3 mgd. For purposes of this example, the average flow is used as the desired flow rate out of the equalization basin. The diurnal peak and minimum flow rate for this example are 1.7 and 0.45 times the average, respectively.

The next step involves the actual construction of the hydrograph. The hydrograph for this example is shown on Figure 3-3. The inflow mass diagram is plotted first. To do this, the hourly diurnal flows are converted to equivalent hourly volumes, and accumulated over the

FIGURE 3.7

RAW WASTEWATER FLOW AND BOD VARIATION
BEFORE EQUALIZATION

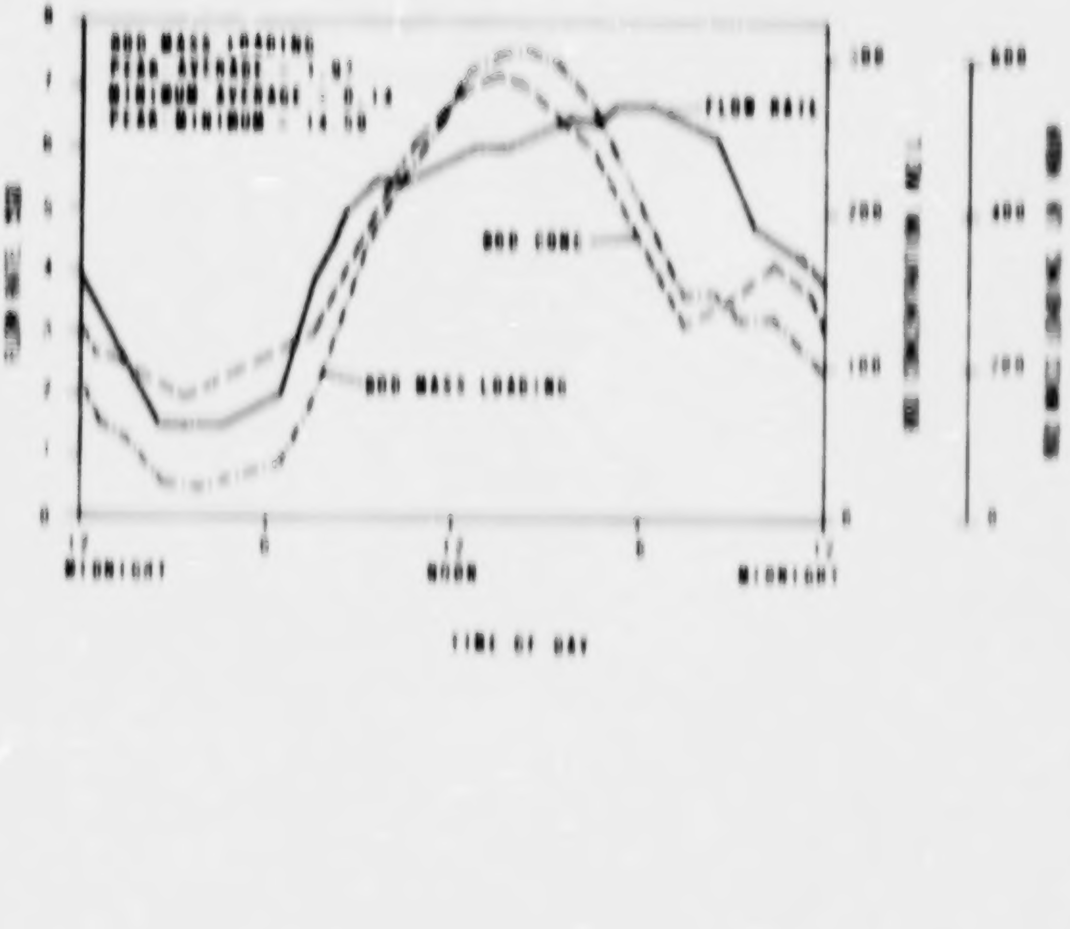
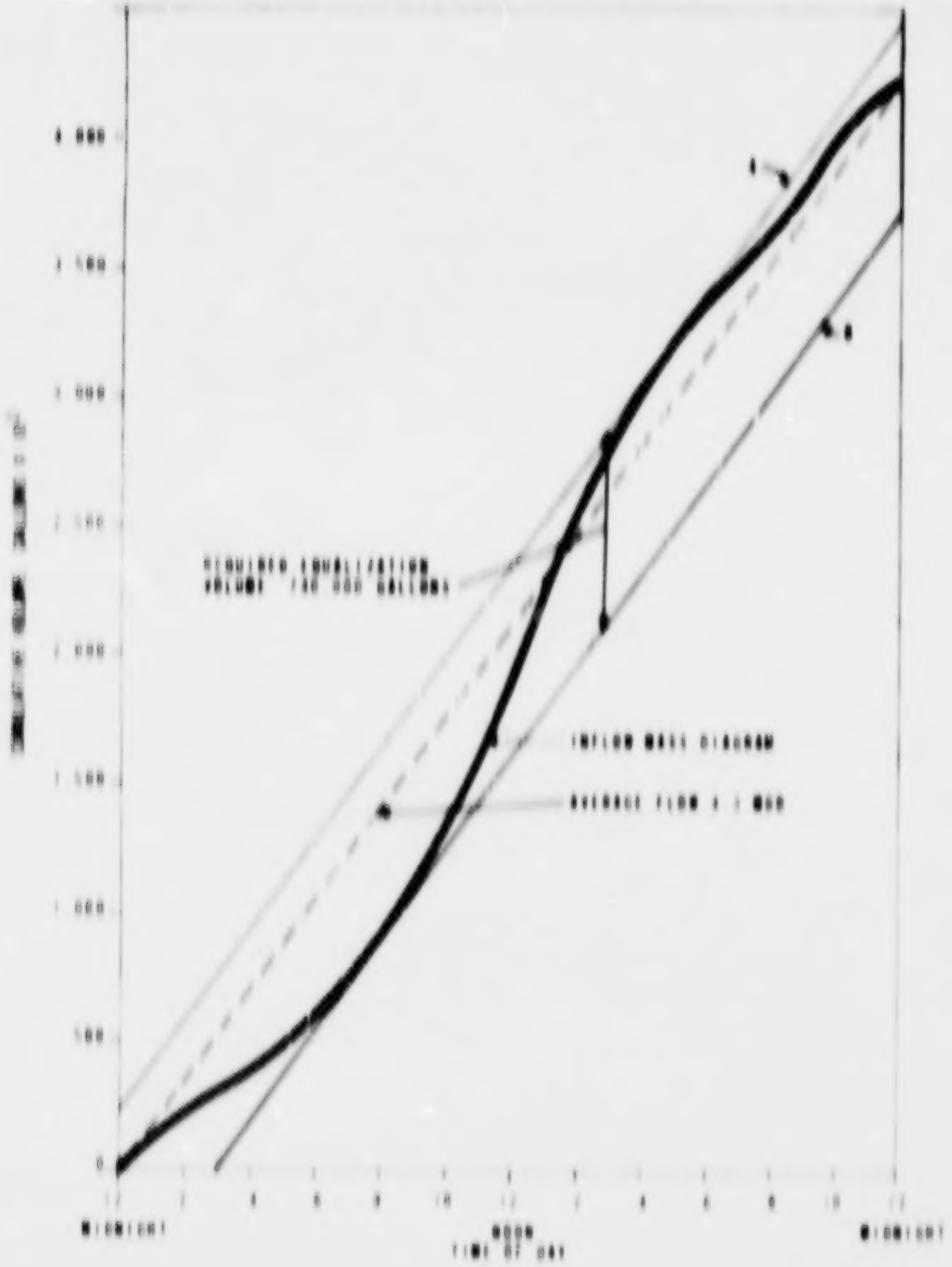


FIGURE 3.3
HYDROGRAPH FOR TYPICAL DIURNAL FLOW



24-hour day. A line is then drawn from the origin to the end point on the inflow mass diagram. The slope of this line actually represents the average flow for the day.

Enough tank volume must be provided to accumulate flows above the equalized flow rate. This normally requires a volume equivalent to 10 to 20 percent of the average daily dry weather flow. To determine this volume, the inflow mass diagram must be enveloped with two lines parallel to the average flow line and tangent to the extremities of the inflow mass diagram. These are shown as lines A and B on Figure 3-3. The required volume is represented by the vertical distance between these two lines. In this illustration, the required volume for equalization is 740,000 gallons, which represents approximately 17 percent of the average daily flow.

The physical interpretation of the hydrograph is simple. At 8:00 AM, the equalization basin is empty, as signified by the tangency of the inflow mass diagram with the bottom diagonal. At this point, plant flow begins to exceed the average flow rate and the tank begins to fill. This is represented by the divergence of the inflow mass diagram and the bottom diagonal. At 5:00 PM, the basin is full, as shown by the tangency of the inflow mass diagram with the top diagonal. Finally, the tank is drawn down from 5:00 PM to 8:00 AM on the following day, when the flow is below average.

The actual equalization basin volume must be greater than that obtained with the hydrograph for several reasons, including:

1. Continuous operation of aeration and mixing equipment will not allow complete drawdown.

2. Volume must be provided to accommodate anticipated concentrated plant recycle streams.

3. Some contingency should be provided for unforeseen changes in diurnal flow.

The final volume selected should include adequate consideration of the conditions listed above and will also depend on the basin geometry. For the example presented herein, a basin volume of approximately one million gallons is adequate.

3.3.2 Impact of Equalization on Diurnal Concentration Variation

At this point, it is appropriate to examine the impact of flow equalization on mass loading and concentrations. As previously mentioned, side-line equalization has a minimal effect on diurnal concentration variations. The following discussion is therefore limited to in-line basins.

An hourly concentration plot for raw wastewater BOD is plotted with the diurnal flow pattern on Figure 4-2. Note that low BOD concentrations occur at night with low flows, and high BOD concentrations occur during the daytime with high flows. This is a typical pattern for dry weather flows and BOD's. Because of this characteristic, the mass loading rate of raw wastewater BOD, shown on Figure 4-3, exhibits even greater fluctuations. If this wastewater is equalized in a one million gallon in-line basin, the equalized flow will exhibit the characteristics shown on Figure 4-4, provided:

1. The basin is designed to provide complete mixing.

2. There is no BOD reduction in the basin.

This damping effect would be similarly beneficial for all concentration variables including SS, nitrogen, phosphorus, and toxic constituents.

On Figure 4-4, the changes in BOD concentration are most pronounced during periods of minimum wastewater volume in the equalization tank. If desired, increased damping can be achieved by increasing the active volume of the tank, i.e., the volume in excess of that obtained from the hydrograph.

4.3.3 Basin Construction

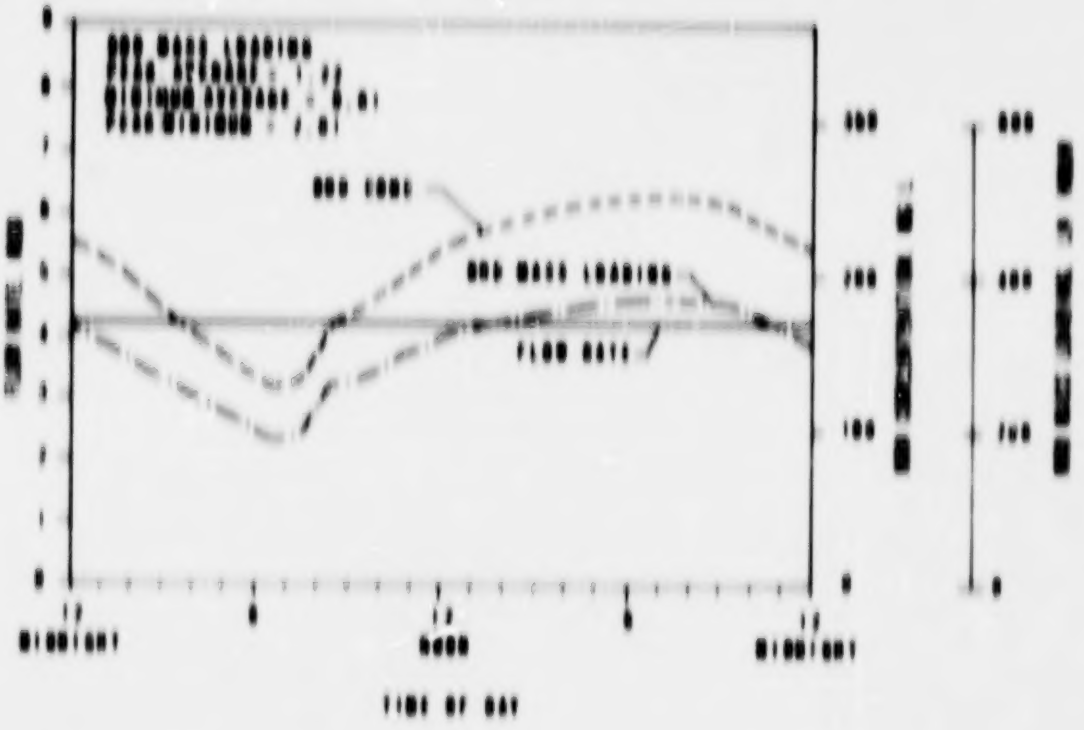
Equalization basins can be provided through the construction of new facilities or through the modification of existing facilities of sufficient volume. Equalization may be implemented with relative ease in an upgrading plan that calls for the abandonment of existing tankage. Facilities which may be suitable for conversion to equalization basins include aeration tanks, clarifiers, digesters and sludge lagoons.

New basins may be constructed of earth, concrete or steel. Earthen basins are generally the least expensive. They can normally be constructed with side slope varying between 3:1 and 2:1 horizontal to vertical, depending on the type of lining used. To prevent embankment failure in areas of high groundwater, drainage facilities should be provided for groundwater control. In large basins where a combination of aerator action and wind forces may cause the formation of large waves, precautions should be taken in design to prevent erosion. It is also customary to provide a concrete pad directly under the equalization basin aerator or mixer. The top of the dikes should be wide enough to ensure a stable embankment. For economy of construction, the top width of the dike should be sufficient to accommodate mechanical compaction equipment.

In-line basins should be designed to achieve complete mixing in order to optimize concentration damping. Elongated tank design enhances plug flow and should be avoided. Inlet and outlet configurations should be designed to prevent short-circuiting. Designs which discharge influent flow as close as possible to the basin mixers are preferred.

FIGURE 3-4

RAW WASTEWATER FLOW AND BOD VARIATION AFTER EQUALIZATION



To continue the previous illustration, an earthen basin has been selected for the equalization facility. A square plan has been chosen to effect optimum mixing. A section view of the basin with appropriate dimensions is shown on Figure 3-6. The volume requirement computed from the hydrograph is provided in the upper eight feet. Note that the minimum required operating depth lies above the minimum allowable aerator operating level.

3.3.4 Air and Mixing Requirements

The successful operation of both in-line and side-line basins requires proper mixing and aeration. Mixing equipment should be designed to blend the contents of the tank, and to prevent deposition of solids in the basin. To minimize mixing requirements, grit removal facilities should precede equalization basins wherever possible. Aeration is required to prevent the wastewater from becoming septic. Mixing requirements for blending a municipal wastewater having a SS concentration of approximately 200 mg/l range from 0.02 to 0.04 hp/1,000 gallons of storage. To maintain aerobic conditions, air should be supplied at a rate of 1.25 to 2.0 cfm/1,000 gallons of storage (9).

Mechanical aerators are one method of providing both mixing and aeration. The oxygen transfer capabilities of mechanical aerators operating in tap water under standard conditions vary from 3 to 4 lb O_2/hp-hr. Baffling may be necessary to ensure proper mixing, particularly with a circular tank configuration. Minimum operating levels for floating aerators generally exceed five feet, and vary with the horsepower and design of the unit. Low level shutoff controls should be provided to protect the unit. The horsepower requirements to prevent deposition of solids in the basin may greatly exceed the horsepower needed for blending and oxygen transfer. In such cases, it may be more economical to install mixing equipment to keep the solids in suspension and furnish the air requirements through a diffused air system, or by mounting a surface aerator blade on the mixer.

It should be cautioned that other factors including maximum operating depth and basin configuration affect the size, type, quantity and placement of the aeration equipment. In all cases, the manufacturers should be consulted.

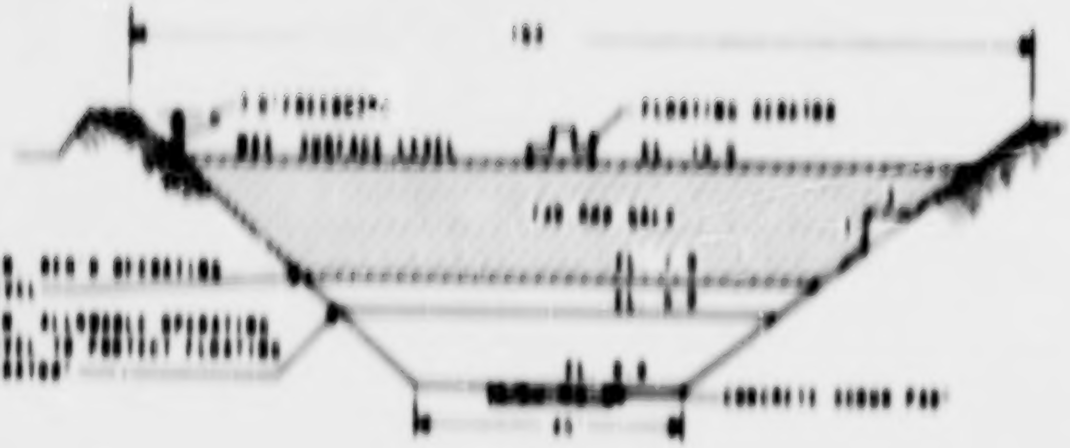
3.3.5 Pump and Pump Control Systems

Flow equalization imposes an additional head requirement within the treatment plant. As a minimum, this head is equal to the sum of the dynamic losses and the normal surface level variation. Additional head may be required if the basin is to be dewatered to a downstream location. It may be possible to dewater the basin upstream (e.g., ahead of raw wastewater pumps) by gravity.

Normally, the head requirement cannot be fulfilled by gravity, thereby requiring pumping facilities. The pumping may precede or follow equalization. In some cases pumping of both

FIGURE 3.5

EARTHEN EQUALIZATION BASIN



FLOATING AERATOR
MAX. SURFACE LEVEL
MIN. OPERATING LEVEL
CONCRETE SCOUR PAD

VOLUMES

EL. [illegible] TO EL. [illegible] APPROXIMATELY [illegible] GAL

EL. [illegible] TO EL. [illegible] APPROXIMATELY [illegible] GAL

TOTAL VOLUME 1,000,000 GAL

*THESE DIMENSIONS WILL VARY WITH AERATOR DESIGN AND MANUFACTURER raw and equalized flows will be required. Influent pumping will require larger capacity pumps to satisfy diurnal peaks.

Gravity discharge from equalization will require an automatically controlled flow regulating device.

A flow measuring device is required downstream of the basin to monitor the equalized flow. Instrumentation should be provided to control the preselected equalization rate by automatic adjustment of the basin effluent pumps or flow regulating device.

3.3.6 Miscellaneous Considerations

The following features are considered to be desirable for the design of the equalization facility:

1. Equalization should be preceded if possible with screening and grit removal to prevent grit deposition and rag fouling of equipment in the basin.
2. Surface aerators should be fitted with legs to support and protect the units when the tank is dewatered.
3. Facilities should be provided to flush solids and grease accumulations from the basin walls.
4. A high-water level takeoff should be provided for withdrawing floating material and foam.
5. An emergency overflow should be provided.

3.4 Costs

The development of alternatives for any plant upgrading program should include at least one which incorporates flow equalization. In all cases, the added cost of flow equalization must be measured against (1) the savings in cost of modifying downstream processes to accept diurnal variations and (2) the improved performance that can be achieved by operating downstream processes under relatively constant loading conditions.

The cost of flow equalization will vary considerably from one application to another, depending on the basin size, construction selected, mixing and aeration requirements, availability of land, location of facility, and pumping requirements. Some judgment must be made on the distribution of pumping costs. Pumping costs for an equalization basin used to upgrade existing facilities should be charged to the basin.

Capital costs for equalization facilities have been estimated by Smith, et al. (10) and are listed in Table 3-2. The costs for earthen basins include plastic liner and floating mechanical aerators. The costs for the concrete basins include diffused aeration facilities. Pumping costs are based on the construction of a separate equalization basin effluent pumping station. The costs were developed in conjunction with activated sludge treatment system designs and therefore include a proportional amount of the engineering fees and interest during construction.

TABLE 3-2
COST OF EQUALIZATION FACILITIES
(EPA INDEX 175)

Plant Size mgd	Basin Size mil gal	Earthen Basin With Pumping	Earthen Basin Without Pumping	Concrete Basin With Pumping	Concrete Basin Without Pumping
1	0.32	$124,000	$ 72,000	$175,000	$134,000
5	0.80	170,000	84,000	333,050	247,000
10	2.40	310,000	134,000	775,000	595,000

The construction cost for the earthen equalization basin on Figure 3-5 is estimated at $80,000. The cost includes excavation, plastic liner, sand subbase, concrete scour pad, dike fill, underdrain and a 40-hp floating aerator. The costs do not provide for pumping costs, land costs, engineering and legal fees, nor interest during construction.

3.5 Performance and Case Histories

Little full-scale operating data are currently available to compare the performance of wastewater treatment plants with and without flow equalization. However, an increasing number of plant designs are incorporating the use of equalization facilities for upgrading existing plants and construction of new plants. The following case histories are presented as examples of equalization basin design.

3.5.1 Ypsilanti Township, Michigan

A flow equalization project at the Ypsilanti Township Sewage Treatment Plant is currently under way. The treatment facility consists of two adjacent activated sludge plants recently upgraded from 7.0 mgd to treat a total flow of 9.0 mgd. Two 350,000-gallon digesters have been converted to equalization tanks. Data will be collected over a two-year study period for each plant. The flow will be equalized in one plant the first year while background data is collected for the remaining plant. The situation will be reversed the second year, with the flow being equalized to the second plant while unequalized flow performance data are collected on the first plant. Comparison of these data will be made to determine the beneficial effects of flow equalization on each plant.

4.5.9 Fond du Lac, Wisconsin

This case illustrates a situation in which only a portion of the flow is equalized. The City of Fond du Lac, Wisconsin, presently employs a single-stage trickling filter plant to treat combined municipal-industrial wastes. Placed in operation in 1950, the plant was designed to treat an ultimate dry weather flow of 8 mgd and a BOD loading of 12,500 lb/day. The facility is presently treating an average of 7.1 mgd with a BOD loading of 24,000 lb/day, and hence is organically overloaded. This condition is aggravated by the fact that the waste discharges from a major industrial contributor (a tannery) are presently concentrated during daylight hours. The tannery discharges wastes to the treatment plant via a separate force main. It accounts for about 35 percent of the BOD and 50 percent of the SS into the plant, and about 15 percent of the influent flow, resulting in a widely fluctuating BOD and SS diurnal load profile.

The wide fluctuations in organic loading are resulting in reduced performance of the trickling filters. This, in conjunction with the advent of more stringent treatment standards, has rendered this facility inadequate. Plans are presently under way to upgrade the treatment plant.

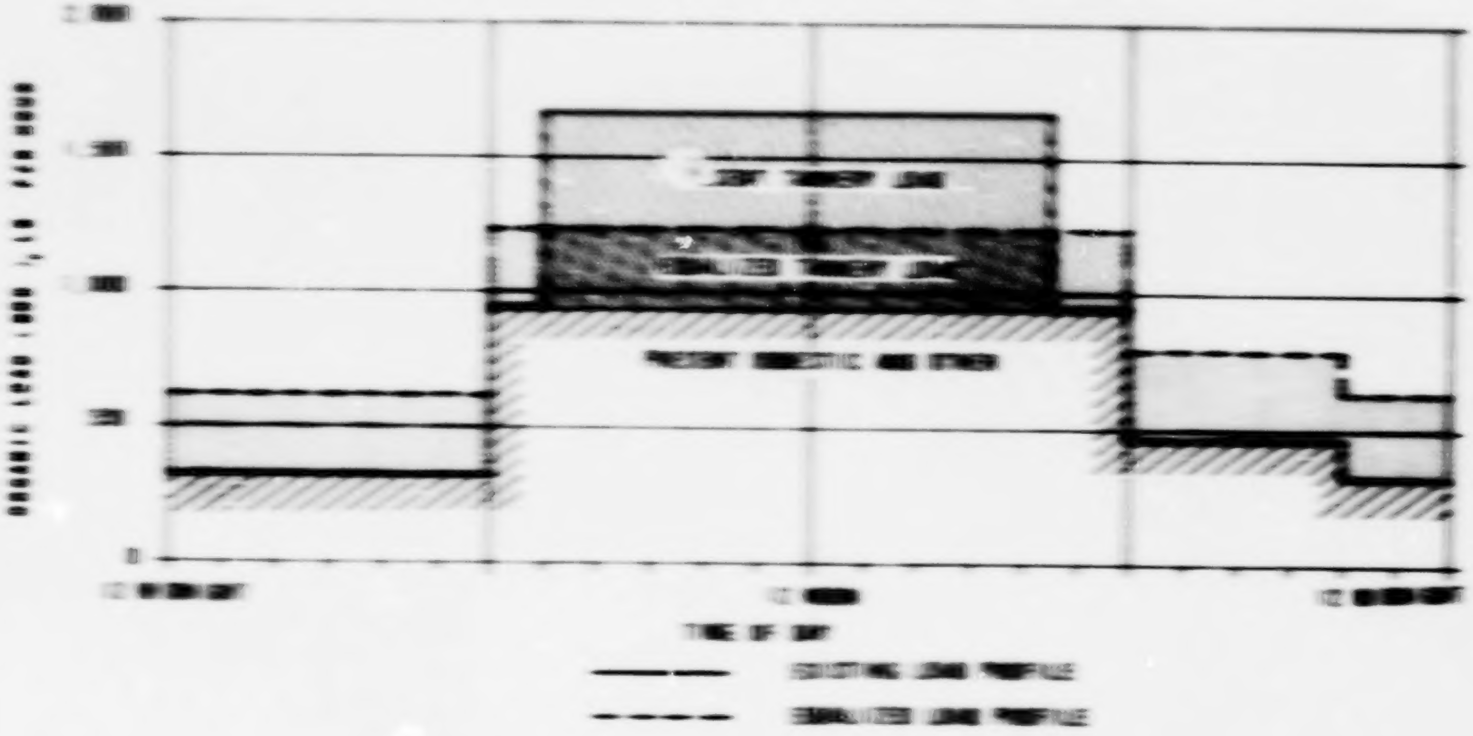
This case represents an ideal situation for employing partial equalization in the upgrading scheme. The volume of the wastes from the tannery is relatively small compared to the total volume of flow received at the plant, whereas the organic contribution is significant. Therefore, a relatively small volume equalization tank is all that is required to attain effective organic load equalization. In addition, because the tannery discharges to the treatment plant via a separate force main, equalization may be accomplished at the treatment plant site. The effect of equalizing the tannery flow over 24 hours is illustrated on Figure 4-6.

Located at the plant site are six abandoned square anaerobic digesters, each measuring 50 feet by 50 feet by 17.5 feet deep. Four of the units have fixed covers and two have floating covers. The utilization of these tanks for equalizing the tannery flow was investigated. The investigations indicate that the four fixed cover tanks would be adequate for equalization for all but a few days each year when the use of the two additional tanks would be necessary because of high flows or maintenance.

The conversion of the abandoned digesters to equalization tanks entails complete modification of the four fixed covered tanks and only minimal modification of the two tanks which have floating covers. The four fixed covered tanks would each require the

FIGURE 3-6

EFFECT OF TANNERY FLOW EQUALIZATION



installation of a mechanical mixer to maintain solids in suspension, including structural modifications in order to support the mixers. A ventilation system would be required for the covered tanks to ensure the safety of plant personnel who may enter the tanks for purposes of inspection or maintenance. Minor structural repairs and waterproofing of all six tanks would be necessary to ensure their structural integrity and watertightness. The two floating covers would be removed and the pipe gallery would be converted to a pump station.

The cost for converting these units to equalization tanks is estimated at approximately $440,000. This cost includes process pumping equipment and piping, four mechanical mixers, tank ventilation system, instrumentation, electrical work, structural renovations and alterations, and engineering fees.

At present, additional studies are under way to evaluate the feasibility of equalization of tannery wastes at the tannery in lieu of equalizing these wastes at the plant site.

3.5.3 Walled Lake – Novi, Michigan

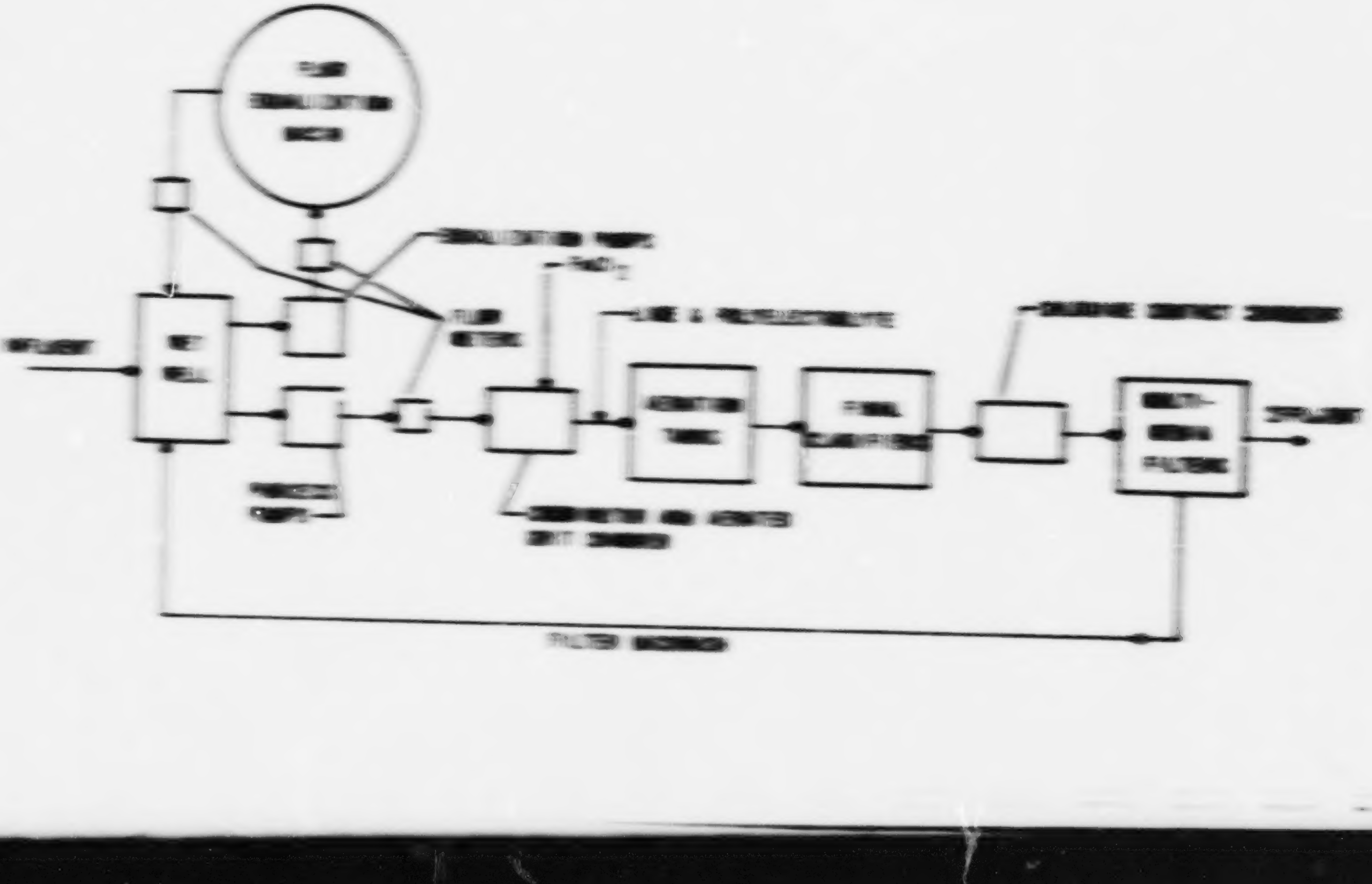
The Walled Lake-Novi Wastewater Treatment Plant is a new 2.1 mgd facility employing side-line flow equalization. The treatment plant was placed into operation in 1971. It was designed to meet stringent effluent quality standards, including (1) a summertime monthly average BOD_5 of 8 mg/l, (2) a wintertime monthly average BOD_5 of 15 mg/l, and (3) 10 mg/l of SS. The facility utilizes the activated sludge process followed by multimedia tertiary filters. Ferrous chloride and lime are added ahead of aeration for phosphorus removal. Sludge is processed by aerobic digestion, and dewatered on sludge drying beds. A schematic diagram of this facility is shown on Figure 3-7.

A major factor in the decision to employ flow equalization was the desire to load the tertiary filters at a constant rate. The equalization facility consists of a 315,000-gallon concrete tank which is equivalent in volume to 15 percent of the design flow. The tank is 15 feet deep and 60 feet in diameter. Aeration and mixing are provided by a diffused air system with a capacity of 2 cfm/1,000 gallons of storage. Chlorination is provided for odor control. A sludge scraper is installed to prevent consolidation of the sludge.

The equalization facility is operated as follows (11). The process pumping rate is preset on the pump controller to deliver the estimated average flow to the treatment processes. The flow delivered by these pumps is monitored by a flowmeter which automatically adjusts the speed of the pumps to maintain the average flow rate. When the raw wastewater flow to the wet well exceeds the preset average, the wet well level rises, thereby actuating variable speed equalization pumps which deliver the excess flow to the equalization basin. When the inflow to the wet well is less than the average, the wet well level falls and an automatic equalization basin effluent control valve opens. The valve releases enough wastewater to the wet well to reestablish the average flow rate through the plant. Since this is a new plant as opposed to

FIGURE 3.7

WALLED LAKE-NOVI WASTEWATER TREATMENT PLANT


FLOW EQUALIZATION BASIN
EQUALIZATION PUMPS
$FeCl_3$
FLOW METERS
LIME & POLYELECTROLYTE
CHLORINE CONTACT CHAMBER
INFLUENT
WET WELL
AERATION TANKS
FINAL CLARIFIERS
MULTI-MEDIA FILTERS
EFFLUENT
FILTER BACKWASH

an upgraded plant, no comparative data exist. However, the treatment facility is typically producing a highly treated effluent, with BOD and SS less than 4 mg/l and 5 mg/l, respectively (9).

3.5.4 Novi Interceptor Retention Basin, Oakland County, Michigan

This case (12) illustrates the utilization of an equalization basin within the wastewater collection system.

A portion of the wastewater collection system for the City of Novi, Michigan, discharges to the existing Wayne County Rouge Valley Interceptor System. Due to the existing connected load on the Wayne County system, Novi's wastewater discharge to the interceptor system is limited to a maximum flow rate of 4 cfs. This rate was matched by the existing maximum diurnal flows from the city. In order that additional population could be served, it was decided to equalize wastewater flows to the interceptor system. By discharging to the interceptor continuously at an average rate of flow, the total wastewater flows from the City of Novi to the Wayne County Rouge Valley Interceptor system could be increased by a factor of 2.6.

An 87,000-cu ft concrete basin was constructed for equalizing flows. The tank has a diameter of 92 feet and a depth of 10.5 feet. Aeration and mixing are provided by a diffused air system with a capacity to deliver 2 cfm/1,000 gallons of storage.

A manhole located upstream of the equalization basin intercepts the flow in the existing Novi trunk sewer. The intercepted wastewater flows into a weir structure which allows a maximum of 4 cfs to discharge into the Wayne County system. The wastewater in excess of the preset average overflows into a wet well where it is pumped to the equalization basin. When flows in the interceptor fall below the preset average, a flow control meter generates a signal opening an automatic valve on the effluent line of the basin, allowing stored wastewater to augment the flow.

3.6 References

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CHAPTER 4

TECHNIQUES FOR UPGRADING TRICKLING FILTER PLANTS

4.1 General

In 1973 there were more than 3,500 trickling filter plants in the United States serving over 30 million people. In contrast, there were approximately 4,750 activated sludge plants serving 80 million people (1). In the past, the trickling filter plant has been considered the ideal plant for populations of 2,500 to 10,000.

Several reasons have justified this popularity. One is its economy, not only in first cost, but also in operation; another is its relative simplicity of operation, which does not require as highly skilled operators as activated sludge plants.

Although activated sludge plants provide higher levels of treatment and greater operational flexibility than trickling filter plants, trickling filter plant performance was considered adequate where stream assimilative capacity was relatively large in relation to population. However, increased urbanization and more stringent effluent and water quality standards will require that many existing trickling filter plants be upgraded.

Upgrading of a trickling filter may be required due to hydraulic or organic overloading, higher effluent quality requirements, or both. In general, decreasing hydraulic or organic overloading in existing facilities will not produce a significant increase in BOD and SS removals above the original design values. However, the effluent polishing techniques discussed in Chapter 7 may provide enough additional treatment to obtain an effluent of the desired quality.

4.2 Trickling Filter Processes

Trickling filtration consists of uniform distribution of wastewater over the trickling filter media by a flow distributor. A large portion of the wastewater applied to the filter passes rapidly through it, and the remainder trickles slowly over the surface of the biological slime which forms. BOD removal occurs by biosorption and coagulation from the rapidly moving portion of the flow and by progressive removal of soluble constituents from the more slowly moving portion of the flow.

The quantity of biological slime produced is controlled by the available food, and the growth will increase as the organic load increases until a maximum thickness is reached. This maximum growth is controlled by hydraulic rate, ventilation, type of media, type of organic matter, amount of essential nutrients present and the nature of the particular biological growth.

In the past, trickling filters have been classified as either low (standard), intermediate, or high rate filters depending upon hydraulic and organic loading rates.

4.2.1 Low Rate Trickling Filters

Low rate trickling filters are designed to handle organic loadings of 5 to 20 lb BOD/day/1,000 cu ft and hydraulic loadings of 1 to 4 mgd/acre. In general, low rate filters do not use recirculation to maintain a constant hydraulic loading, but use either suction level controlled pumps or a dosing siphon. Dosing tanks are small, usually with only a 2-minute detention time based on twice the average design flow, so that intermittent dosing is minimized. Even so, at small plants, low nighttime flows may result in intermittent dosing. If the interval between dosing is greater than one or two hours, the efficiency of the process will deteriorate since the character of the biological slime will be altered due to lack of moisture. Under normal conditions, the BOD removal efficiency of a low rate filter and secondary clarifier may average 75 to 85 percent. By the addition of recirculation during periods of low flows to keep the filter wet, it is possible to increase overall plant efficiency (2).

Low rate filters are normally constructed using a 5 to 10 foot depth of stone media. In most low rate filters, only the top 2 to 4 feet of the filter media have appreciable biological slime. As a result, the lower portions of the filter may be populated by autotrophic nitrifying bacteria which oxidize ammonia nitrogen to nitrite and nitrate forms. If the nitrifying population is sufficiently well established, and if climatic conditions and wastewater characteristics are favorable, a well operated low rate filter, in addition to providing good BOD removal, can produce a highly nitrified effluent. The benefits of well nitrified effluents for reduction of total oxygen demand in receiving waters are being increasingly utilized in the formulation of effluent standards.

4.2.2 Intermediate Rate Trickling Filters

Intermediate rate trickling filters are generally designed to treat hydraulic loadings of 4 to 10 mgd/acre, including recirculation, and organic loadings, excluding recirculation, range from 15 to 30 lb BOD/day/1,000 cu ft. In the past, there have been some cases where the organic loading in the intermediate range stimulated considerable biological filter growth and the rate of hydraulic loading was not sufficient to eliminate clogging of the trickling filter media (2). This clogging situation can be remedied somewhat by utilizing relatively large stone, 3 to 4 inches in diameter. However, many plants operate in the intermediate loading range without reported operational problems (2) (3). In practice, many high rate filters will operate in the intermediate range during the early low flow period of their operating lifetime.

4.2.3 High-Rate Trickling Filters

High-rate trickling filters have hydraulic loadings of 10 to 30 mgd/acre, including recirculation, and organic loadings of 30 to 60 lb BOD/day/1,000 cu ft, excluding recirculation. Media depths of 3 to 6 feet are commonly employed. In all high-rate filters, recirculation is used to maintain a relatively constant hydraulic loading. The higher organic loadings in high-rate filters preclude the development of nitrifying bacteria in the lower section of the filter. Hence, these plants will seldom exhibit any nitrification, and will generally not perform as well as low-rate filters.

4.2.4 Super-Rate Trickling Filters

Super-rate trickling filters have evolved as a result of the development of various types of synthetic media. Past experience has indicated that hydraulic loadings of 150 mgd/acre and higher, including recirculation, may be accommodated in super-rate trickling filters. The major application of super-rate filters has been for high-strength wastes and as roughing units. Synthetic media filters, because of their high surface area per unit volume, can perform as well as high-rate filters at volumetric BOD loadings of about 50 to 100 lb/1,000 cu ft and hydraulic loadings of 0.5 to 1.5 gpm/sq ft. A discussion of synthetic media characteristics is presented in Section 4.4.4.

4.2.5 Trickling Filter Performance Data

Performance data for a number of trickling filters under various operating conditions are presented in Table 4-1. These data emphasize the wide range of effluent quality that can be expected depending upon design and operating conditions.

4.3 Trickling Filter Design Considerations

Trickling filters traditionally have been designed using one of several equations developed over the years from trickling filter plant operating data. The equations incorporate different combinations of the many variables that most affect trickling filter efficiency. Unfortunately, no equation has yet been developed which reflects the actual performance of filters, due to the complex interrelationship of the many variables involved. Since it is difficult to accurately predict trickling filter performance based on these equations, they have limited utility in trickling filter plant design. Wherever possible, it is recommended that treatability or pilot plant studies be used to verify performance predictions based upon the various design equations.

4.3.1 Trickling Filter Design Equations

The trickling filter design equations which are in general use are those published by the National Research Council (8), Ten-State Standards (9), Velz (10), Galler and Gotaas (11)

TABLE 6-1

OPERATING DATA FOR SINGLE-STAGE TRICKLING FILTER PLANTS AT LOW, INTERMEDIATE, AND HIGH LOADING RATES

Plant Location	Influent Flow	Filter Media Depth	Plant Influent BOD	Recirculation Rate	Hydraulic Loading	Organic Loading	Final Effluent BOD	Filter and Final Clarifier BOD Removal	Reference
	mgd	ft	mg/l		mgd/acre	lb BOD/day/1,000 cu ft	mg/l	percent	
Low-Rate									
Aurora, Illinois	7.8	6	167	0	2.1	4.4	16	90	4
Dayton, Ohio	38.5	7.5	157	0	3.5	12	33	76	4
Durham, N. Carolina	3.8	7	225	0	1.9	11	60	73	4
Madison, Wisconsin	4.8	10	288	0	2.8	6.4	19	76	4
Richardson, Texas	1.5	6.5	166	0	3.9	11.1[a]	29	85	5
Intermediate-Rate									
Plainfield, New Jersey	4.3	6	129	0.6	2.8	15	13	83	4
Great Neck, New York	0.5	6	167	3.0	7.8	29	29	83	6
High-Rate									
Oklahoma City, Oklahoma	16.2	6	462	3.0	16.2	78	98	78	6
Fremont, Ohio	3.0	3.3	138	1.5	19.0	41	25	78	6
Storm Lake, Iowa	2.6	8	600	2.3	28.5	62	96	86	6
Richland, Washington	2.2	4.5	173	2.0	19.6	38	29	85	6
Merced, California	0.6	3.2	290	6.8	26.0	75	25	87	6
Chapel Hill, N. Carolina	1.4	3.25	166	2.0	16.5	39	48	82	7

[a] Computed based on assumed 35 percent BOD removal in primary clarifiers.

Schulze (12) and Eckenfelder (13) (14). The National Research Council, Velz, Eckenfelder, and Galler and Gotaas equations are presented and discussed in the following sections.

4.3.1.1 National Research Council (NRC) Equations

The NRC formulation was the result of an extensive analysis of operational records from stone media trickling filter plants serving military installations. Based on data analyses, the NRC recommended the following equations for predicting the performance of stone media trickling filters:

First or Single Stage:

$$E_1 = \frac{100}{1 + 0.0085\left(\frac{W}{VF}\right)^{1/2}}$$

Second Stage:

$$E_2 = \frac{100}{1 + \frac{0.0085}{1 - E_1}\left(\frac{W'}{VF}\right)^{1/2}}$$

where:

E_1 = Percent BOD removal efficiency through the first stage filter and clarifier

W = BOD loading (lb/day) to the first or single stage filter, not including recycle

V = Volume of the particular filter stage in acre feet

F = Recirculation factor for a particular stage; $(1 + R)/(1 + 0.1R)^2$
R = Recirculation ratio = recirculated flow/plant influent flow

E_2 = Percent BOD removal efficiency through the second stage filter and clarifier

W' = BOD loading (lb/day) to the second stage filter, not including recycle

Some of the limitations of the NRC equations are:

1. Military wastewater is characteristically more concentrated than average domestic wastewaters. Consequently, higher percentage removal efficiencies per unit of volume were easier to obtain.

2. The effect of temperature on trickling filter performance is not considered.

3. NRC equations indicate that organic loading has a greater influence on filter efficiency than hydraulic loading. This is probably because of the concentrated nature of the wastewaters.

4. Applicability is limited to concentrated domestic wastewaters because no factor is included to account for differing treatability rates.

5. The equation for second-stage filters is based on the existence of intermediate clarifiers following the first-stage filters.

A comparative plot of trickling filter operating data with the predicted value using the first or single-stage NRC equation is shown on Figure 4-1. It is evident from Figure 4-1 that the use of the NRC equation may result in substantial deviation from the actual performance of a trickling filter.

4.1.1.2 Velz Equation

In 1948, Velz proposed the first major formulation delineating a fundamental law as contrasted to previous attempts based on data analysis. The Velz equation relates the BOD remaining at depth D as follows:

$$\frac{L_D}{L} = 10^{-KD}$$

where:

L = Total removable BOD, mg/l
L_D = Removable BOD at depth D, mg/l
D = Filter depth, feet
K = Constant

Removable BOD in the Velz equation is defined as the maximum fraction of applied BOD removed at a specific hydraulic loading range. When recirculation is used, the total applied BOD, L_a after dilution by recirculation, R, may be determined as follows:

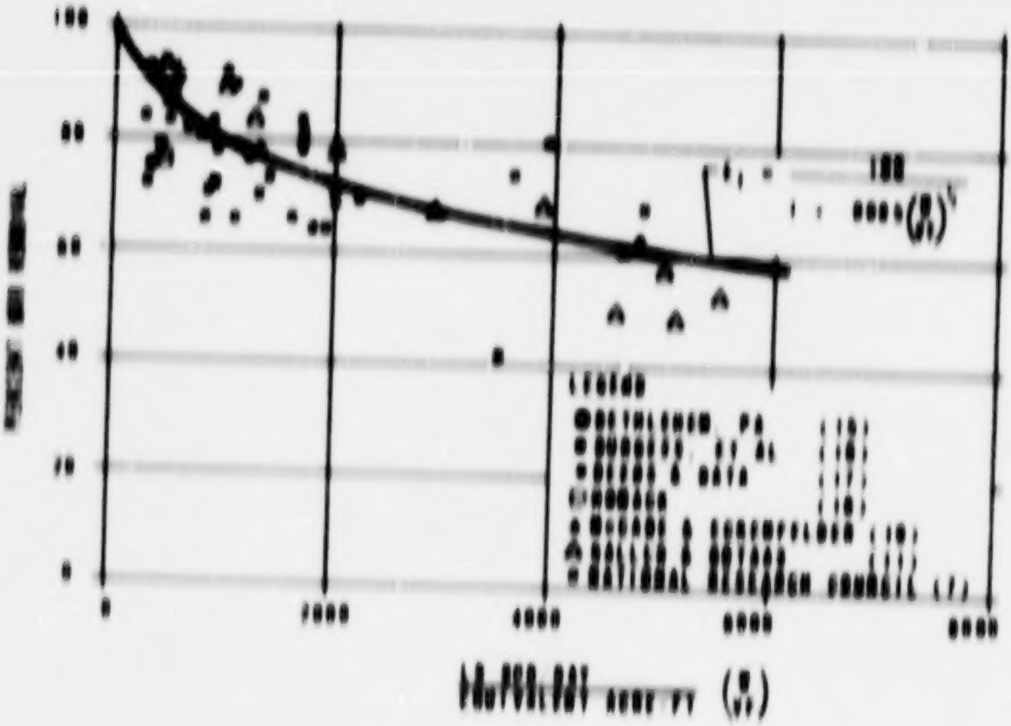
$$L_a = \frac{L_i + RL_e}{1 + R}$$

where:

L_i = Total influent BOD, not including recirculation, mg/l
L_e = Total effluent BOD, mg/l

FIGURE 4.1

COMPARISON OF TRICKLING FILTER OPERATING DATA WITH NRC EQUATION



4.4.1.3 Eckenfelder Equation

Eckenfelder expanded the Velz equation to include other factors. Schulze (12) had postulated that the time of liquid contact with the biological slime is directly proportional to the filter depth and inversely proportional to the hydraulic loading rate. Combining the time of contact with Velz's first order equation for BOD removal and including the effect of changes in filter depth on the BOD removal per unit of depth, Eckenfelder proposed the following equation:

$$\frac{L_e}{L_a} = \frac{1}{1 + \frac{2.5\,D^{0.67}}{\left(\frac{Q}{A}\right)^{0.5}}}$$

L_a = Influent BOD (including recirculation), mg/l
L_e = BOD of unsettled filter effluent, mg/l

With A in acres, D in feet and Q in mgd

4.4.1.4 Galler and Gotaas Equation

In 1964, the last major effort to forecast the performance of stone filters was attempted by Galler and Gotaas (11) using multiple regression analysis of data from existing plants.

Based on regression analysis, the following equation was developed:

$$L_e = \frac{K\,(Q\,L_i + r\,L_e)^{1.19}}{(Q + r)^{0.78}\,(1 + D)^{0.67}\,a^{0.25}}$$

where

$$K = \frac{0.464\left(\frac{43{,}560}{\pi}\right)}{Q^{0.28}\;T^{0.15}}$$

L_e = Unsettled filter effluent BOD, mg/l
L_i = Filter influent BOD, mg/l
D = Filter depth, feet
Q = Influent flow, mgd
r = Recirculation flow, mgd
a = Filter radius, feet
T = Wastewater temperature, deg C

The Galler and Gotaas equation recognized the effects of recirculation, hydraulic loading, filter depth, and wastewater temperature as being important in understanding the performance of a trickling filter. They further indicated that recirculation improves the performance of a filter, but established a 4:1 ratio as a practical upper limit for recirculation.

4.4.2 Applicability of Various Trickling Filter Design Equations

The design engineer has available several equations for trickling filter designs, and the decision to use one in preference to another is often difficult. The availability of several equations often raises doubts concerning their validity.

An attempt has been made by Hanumanulu (20) to compare the actual performance of a 12-foot deep, stone media trickling filter with that predicted using NRC, Ten-States Standards, Velz, Eckenfelder, and Galler and Gotaas equations. The filter was operated at a constant flow without recycle as well as with a 1:1 recirculation ratio. It was found that the Velz, Ten-States Standards and the NRC equations predicted filter efficiencies that are closer to observed values when operated without recycle, while the Eckenfelder and Galler and Gotaas equations predicted efficiencies closer to observed values for filters operated with recirculation. However, because the filter studied was about double the depth of most filters, the conclusions may have differed from those that would have been observed with a shallow filter.

Ordon (21) calculated the volume of filter media required to achieve specified BOD removals using the NRC, Eckenfelder, and Galler and Gotaas equations. The wastewater flow, BOD, and temperature were assumed as 1 mgd, 100 mg/l, and 20 deg C, respectively. The trickling filter volume calculated by the different equations is shown in Table 4-3.

Inspection of Table 4-3 indicates characteristic trends which the designer should be aware of before using any of these equations. When recirculation was zero, the filter volumes calculated from the NRC and Eckenfelder equations were essentially the same, while the Galler and Gotaas equation gave volumes which were significantly different. However, when recirculation was considered, the NRC design volumes were generally quite conservative, while the volumes calculated by Eckenfelder and Galler and Gotaas equations were more nearly the same. In general, the NRC equations would seem to apply when recirculation is not considered, when seasonal temperature differentials are minor, and when the wastewater load is highly variable and of high strength.

4.4.3 Laboratory and Pilot Plant Treatability Studies

As previously indicated, treatability or pilot plant studies are advantageous in verifying the performance predicted by design equations. The use of treatability studies for design of trickling filters has been limited due to the lack of suitable laboratory scale testing methods.

TABLE 4.2

TRICKLING FILTER VOLUMES FOR VARIOUS ORGANIC REMOVALS AS CALCULATED BY DIFFERENT DESIGN EQUATIONS (ALL VOLUMES IN THOUSANDS OF CUBIC FEET)

Recirculation	50% BOD Removal			60% BOD Removal			70% BOD Removal			[illegible] BOD Removal			80% BOD Removal			90% BOD Removal		
Ratio	NRC[1]	Eck.[2]	G&G[3]	NRC	Eck.	G&G	NRC	Eck.	G&G	NRC	Eck.	G&G	NRC	Eck.	G&G	NRC	Eck.	G&G
0	[illegible]	[illegible]	[illegible]	[illegible]	[illegible]	[illegible]	[illegible]	[illegible]	[illegible]	[illegible]	[illegible]	[illegible]	[illegible]	[illegible]	[illegible]	[illegible]	[illegible]	
1	[illegible]	[illegible]	[illegible]	[illegible]	[illegible]	[illegible]	[illegible]	[illegible]	[illegible]	[illegible]	[illegible]	[illegible]	[illegible]	[illegible]	[illegible]	[illegible]	[illegible]	
2	[illegible]	[illegible]	[illegible]	[illegible]	[illegible]	[illegible]	[illegible]	[illegible]	[illegible]	[illegible]	[illegible]	[illegible]	[illegible]	[illegible]	[illegible]	[illegible]	[illegible]	[illegible]
3	[illegible]	[illegible]	[illegible]	[illegible]	[illegible]	[illegible]	[illegible]	[illegible]	[illegible]	[illegible]	[illegible]	[illegible]	[illegible]	[illegible]	[illegible]	[illegible]	[illegible]	[illegible]
4	[illegible]	[illegible]	[illegible]	[illegible]	[illegible]	[illegible]	[illegible]	[illegible]	[illegible]	[illegible]	[illegible]	[illegible]	[illegible]	[illegible]	[illegible]	[illegible]	[illegible]	[illegible]
5	[illegible]	[illegible]	[illegible]	[illegible]	[illegible]	[illegible]	[illegible]	[illegible]	[illegible]	[illegible]	[illegible]	[illegible]	[illegible]	[illegible]	[illegible]	[illegible]	[illegible]	[illegible]
6		[illegible]	[illegible]		[illegible]	[illegible]		[illegible]	[illegible]		[illegible]	[illegible]		[illegible]			[illegible]	

[1] NRC - National Research Council

[2] Eck. - Eckenfelder

[3] G&G - Galler and Gotaas

Design Conditions:
Filter Influent Flow 1 mgd
Filter Influent BOD 200 mg/l
Wastewater Temperature 20 deg C

and has generally been restricted to synthetic media. Pilot units can be rented from the media manufacturers but will require considerable manpower and funds to obtain the meaningful data needed for design purposes. Treatability studies for evaluation of stone media filter design parameters have not been commonly performed.

However, advances are being made in the development of a practical laboratory-scale piloting facility for both stone and plastic media. Based on the concept of contact time as introduced by Schulze, the trickling filter process may be modeled by using an inclined plane to support biological growth (22).

Wastewater is introduced at a variable rate to the top of the slimed inclined plane. The plane's inclination may be varied to change the contact time. As previously discussed, Schulze's equation relates the contact time to the depth and hydraulic loading, as well as to the physical characteristics of the filter media. BOD removal is then assumed to vary with the following first order removal equation:

$$\frac{L_E}{L_I} = e^{-kt}$$

where:

L_E = BOD of unsettled filter effluent, mg/l
L_I = BOD of filter influent, mg/l
k = Treatability constant
t = Contact time, minutes

The inclined plane method furnishes data on BOD removal, contact time, hydraulic loading and recirculation ratios.

Since the basic purpose of either a laboratory or pilot plant evaluation is to study variables that affect filter performance, any treatability study should be of sufficient duration to adequately describe the following variables as they affect the filter performance:

1. Applied BOD loading
2. Hydraulic loading
3. Recirculation
4. Wastewater temperature

The data thus obtained from treatability studies can be evaluated using the various trickling filter equations previously discussed.

4.4 Trickling Filter Upgrading Considerations

The performance of a trickling filter plant is determined by the complex interrelationship of numerous variables. Thus, in any upgrading situation, the acquisition and analysis of year-round performance data are essential in determining the additional treatment facilities needed to meet more stringent effluent requirements. In addition to those variables that directly affect performance of the trickling filter itself, the performance of the plant as a whole also depends on proper integration of the filter with the other unit processes being utilized. The more important variables to be considered prior to upgrading any trickling filter plant are discussed in the following sections.

4.4.1 Wastewater Characteristics

Municipal wastewaters vary in composition and strength, depending on the relative amounts of industrial wastewater and infiltration present. The rate of BOD removal from a domestic wastewater in a trickling filter generally exceeds the BOD removal rate from an industrial wastewater which has a high percentage of dissolved BOD. This is due to the high percentage of colloids in domestic wastewater, and to the increased ability of the filter to remove this colloidal material. A reasonable explanation for this is that some of these materials are removed by biological flocculation and not by oxidation and synthesis of new cells.

A trickling filter plant that requires upgrading due to industrial loads in excess of those anticipated in the design may be improved by equalization of the industrial flows. Often this is most economically accomplished at the industrial source and is particularly advantageous when the entire industrial load is discharged in an 8- or 16-hour period. Under other conditions, equalization facilities at the treatment plant may be provided as discussed in Chapter 3 to reduce peak loadings on the trickling filter.

4.4.2 Recirculation

The practice of effluent recirculation can be used to improve the efficiency and operation of some stone media trickling filters. For example, it can minimize the operational problems associated with intermittent dosing of low-rate trickling filters. Recirculation ratios of 0.5 to 4.0 have been used in high-rate filters. Galler and Gotaas (11) have demonstrated that a recirculation ratio of greater than 4.0 does not materially increase plant efficiency and is also uneconomical. Normal design practice is to use ratios of 1.0 to 2.0.

Recirculation as applied to synthetic media involves a somewhat different concept than is applied to stone filters. Various types of synthetic filter media have higher minimum wetting rates, i.e., a rate of flow per unit area which will induce a biological slime throughout the depth of the media. This minimum wetting rate typically ranges from 0.5 to 1.0 gpm/sq ft (30 to 60 mgd/acre), depending on the geometric configuration of the media. Therefore, recirculation in synthetic media filters is practiced to maintain the desired wetting rate for a particular medium. Generally, increasing the hydraulic loading substantially above the minimum wetting rate decreases the BOD removal through the filter (23).

There are many possible flow configurations which may be used with a single- or two-stage high-rate trickling filter plant. Some of the more common flow diagrams which have been presented in the Water Pollution Control Federation MOP No. 8, Sewage Treatment Plant Design (3) are shown on Figure 4-9. Decisions regarding the use of any one of these flow configurations in a plant upgrading must be based on its suitability to the existing plant facilities and an examination of the relative economics.

The hydraulic capacity of the trickling filter distribution and underdrain systems also requires investigation where changes to the recirculation pattern are proposed. Significant increases in recirculated flow may exceed the capacity of the distribution mechanism or overload the filter media or underdrain conduits.

4.4.3 Clarifier Capacity

When modifying an existing trickling filter plant, the effect of increased recirculation on both the primary and final clarifiers must be considered. Recycle schemes requiring the recycled flow to pass entirely through the primary and/or secondary clarifiers exert significantly higher load on these units than schemes where the recycle passes through the filter alone. Since studies have shown that direct recirculation of trickling filter effluent is as effective as recycling clarified effluent (24), this is the preferred flow pattern where clarifier loading is otherwise excessive. Recommended loadings for trickling filter clarifiers are presented in Chapter 6.

4.4.4 Trickling Filter Media

The physical properties of various types of trickling filter media are shown in Table 4-3.

The properties shown in Table 4-3 which are of greatest interest are specific surface area and percent void space. Greater surface area permits a larger mass of biological slimes per unit volume, while increased void space allows for higher hydraulic loadings and enhanced oxygen transfer. The ability of synthetic media to handle higher hydraulic and organic loadings is directly attributed to the higher specific surface area and void space of these media compared to stone media and blast furnace slag.

FIGURE 4-2
COMMON FLOW DIAGRAMS FOR SINGLE AND TWO-STAGE HIGH-RATE TRICKLING FILTERS (2)

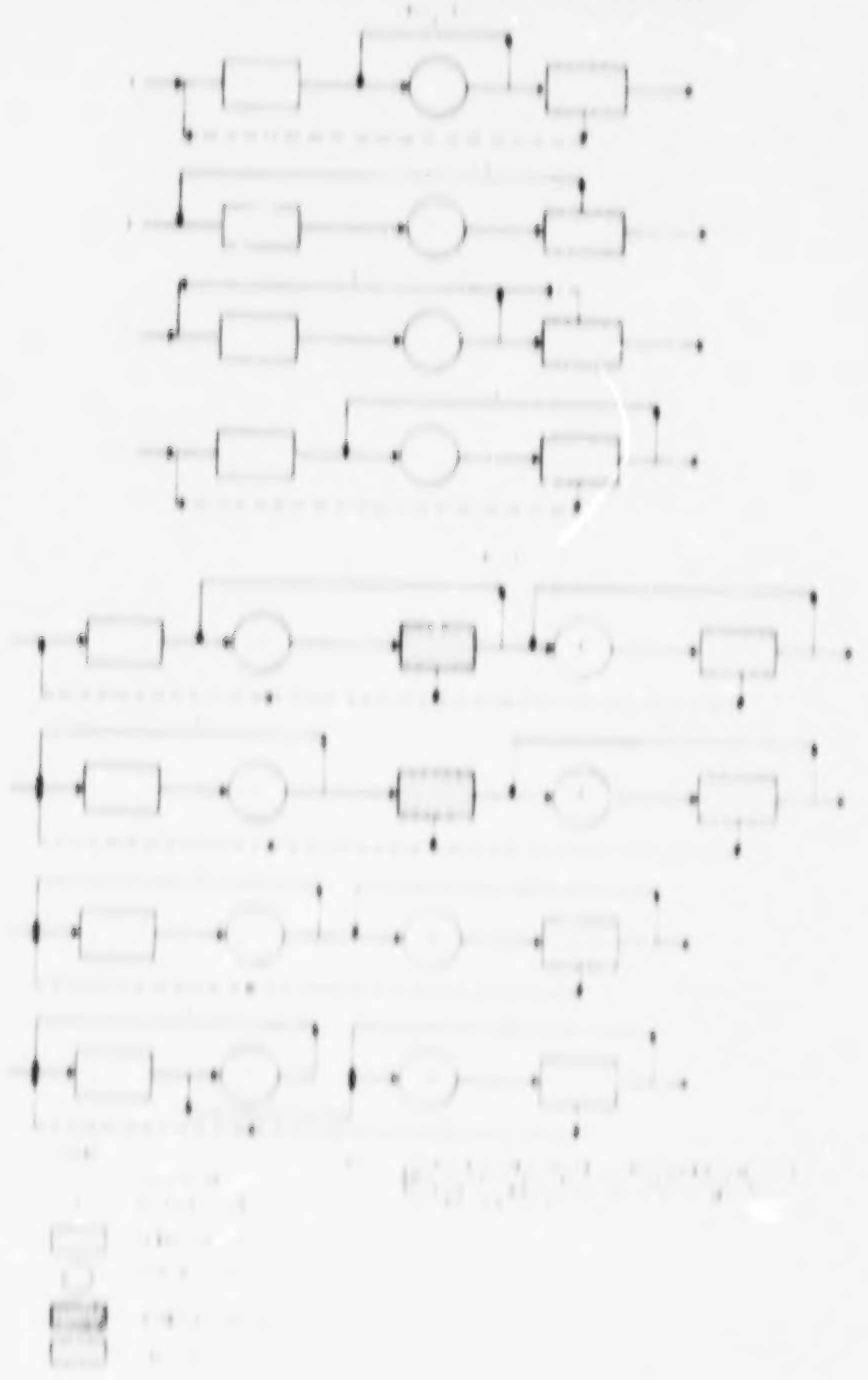


TABLE 4-4
COMPARATIVE PHYSICAL PROPERTIES OF TRICKLING FILTER MEDIA

Media	Nominal Size in	Units per cu ft	Unit Weight lb/cu ft	Specific Surface Area sq ft/cu ft	Void Space percent
Plastic	24 x 24 x 48	2-3	2-6	25-35	94-97
Redwood	47½ x 47½ x 35¾	—	10.3	14	76
Granite	1-3	—	90	19	46
Granite	4	—	—	13	60
Blast Furnace Slag	2-3	50-60	68	20	49

An important consideration in any upgrading involving increased recirculation is the effect of the resulting increased flow through the filter media. In some cases, replacement of an existing filter media with a media having greater void space may be required to avoid ponding due to the higher hydraulic loadings and increased biological growths.

4.4.5 Trickling Filter Depth

Low-rate trickling filters have traditionally been designed with depths ranging from 5 to 10 feet. Because the rate of biological activity is greatest at the surface of a stone media trickling filter and diminishes with depth, high-rate stone media filters are designed with shallower depths of 3 to 6 feet to maximize the rates of treatment per unit volume of filter media.

As previously mentioned, nitrifying bacteria residing in the lower portions of the relatively deep low-rate filters can produce a highly nitrified effluent, while high-rate filters seldom exhibit significant nitrification because of high volumetric BOD loadings. The high loadings favor the growth of heterotrophic carbonaceous organisms rather than autotrophic nitrifying organisms.

Recommended depths for synthetic media are 15 to 40 feet. This is because optimum synthetic media hydraulic wetting rates permit smaller diameter units to be used. Thus, increased depth is required to provide adequate contact time with the wastewater. The low unit weight of synthetic media enhances the construction of much deeper filters than is feasible with stone media.

4.4.6 Ventilation

Proper ventilation of trickling filters is essential to the maintenance of aerobic conditions throughout the filter media. It is recommended that all drains, channels, and pipes be sized such that not more than 50 percent of their cross-sectional area will be submerged at the peak hydraulic loading.

If the trickling filter is constructed on or near grade, provision for ventilation will be less critical than if the topography necessitates construction well below grade. In these latter instances, forced ventilation or ventilation shafts may be necessary.

4.4.7 Temperature of Applied Wastewater

The efficiency of trickling filters is affected by wastewater temperature changes in accordance with the following relationship (25) (26):

$$E_T = E_{20}\,\theta^{T-20}$$

where:

θ = Constant varying from 1.035 to 1.041
E_T = Filter efficiency at temperature T
E_{20} = Filter efficiency at 20 deg C
T = Wastewater temperature, deg C

In northern regions, the effect of both air and wastewater temperatures on trickling filter performance is very evident, and significant deterioration of the plant performance can be expected during the winter months (27). The effect of air temperature is especially pronounced in high-rate filters due to the cooling effect of recirculation, and therefore should be taken into consideration whenever an upgrading involves significant changes in the recirculation pattern. It has been reported that covering of filters in cold climates does not substantially increase the performance because the filter covering does not increase the temperature of the applied wastewater (28).

4.4.8 Sludge Handling

Upgrading secondary treatment facilities usually results in an increase in sludge production. Prior to any trickling filter upgrading, therefore, a thorough evaluation of the effects of the upgrading on the sludge processing system should be made and the results incorporated into the overall plant upgrading plan. Sludge production increases may be particularly significant where chemical precipitation in the secondary clarifiers is practiced.

Humus sludge from trickling filters is commonly returned to the primary clarifier, often as part of the recirculation pattern. Conservative primary clarifier overflow rates must be used in such cases. Where primary tank overflow rates are excessive, separate gravity thickening of combined primary and humus sludge can be employed.

Although sludge processing modifications and/or expansions may be costly, an efficient sludge handling system is crucial to good treatment plant operation. Upgrading considerations for sludge processing facilities are discussed in detail in Chapters 10 through 12.

4.5 Trickling Filter Upgrading Techniques and Design Basis

Upgrading to relieve overloaded conditions, to improve organic removal efficiency, to provide nitrification, and to remove nutrients is covered in the following sections. The choice between the available options will depend on such factors as plant hydraulics, conditions of existing treatment facilities and future effluent requirements. The final selection should maximize the utilization of available clarifiers and minimize modifications in process piping and additional pumping.

4.5.1 Upgrading to Relieve Organic and Hydraulic Overloading

Trickling filter plants may be upgraded to relieve hydraulic and/or organic overloading by any one of the following three general procedures:

1. Upgrading existing single stage filters by the construction of additional trickling filters in parallel with existing units or by conversion of a low rate filter to high rate.

2. Upgrading single stage trickling filters to a two-stage biological system by the addition of second stage trickling filters or an activated sludge system.

3. Upgrading existing two-stage trickling filters to a multiple stage biological system.

Upgrading of single stage filters by the construction of parallel single stage units or by conversion to high rate filters is a straightforward procedure and is not discussed in detail since all of the factors involved have been covered previously. It is emphasized that before a decision is made to adopt this procedure, a careful analysis of plant performance data should be made to determine if effluent criteria can be met.

4.5.1.1 Upgrading a Single-Stage Low-Rate Trickling Filter by Improving Distribution

Upgrading an organically overloaded and hydraulically underloaded single-stage low-rate trickling filter may be accomplished by providing recirculation. This upgrading procedure was used to improve the wastewater treatment plant performance at Pueblo, Colorado (29).

The original secondary treatment facilities consisted of low-rate rock media trickling filter and final clarifier units as shown on Figure 4-3. At the time of design it was determined that a 3-foot filter depth, without recirculation, would provide adequate treatment to meet Colorado State Health Department regulations prohibiting a BOD discharge in excess of 30 mg/l. Table 4-4 shows operating data for the plant. It is apparent from these data that the trickling filters were not operating as well as had been expected. Hydraulic loading was inadequate to keep the distributor arms moving during periods of low flow. As a result, sufficient filter growth could not be maintained throughout the media to adequately treat the relatively high organic loads, especially during the cold weather months when reduced biological activity and freezing occurred. The shallow depth of the filter media aggravated these problems.

The plant was upgraded in 1967 to include recirculation facilities having a capacity of about 80 percent of the average daily flow. A flow diagram of the upgraded plant is also shown on Figure 4-3. Recirculation during periods of low flow provided the necessary flow to keep the distributors revolving and thus prevented loss of the filter growth. The benefits of recirculation were particularly evident in the summer months when the process was not adversely affected by low wastewater temperature. Operating data for the upgraded plant are shown in Table 4-4. Despite an increase of more than 80 percent in filter organic loading, effluent BOD was slightly improved.

The construction costs for the upgrading were estimated to be $129,000 and were allocated as follows:

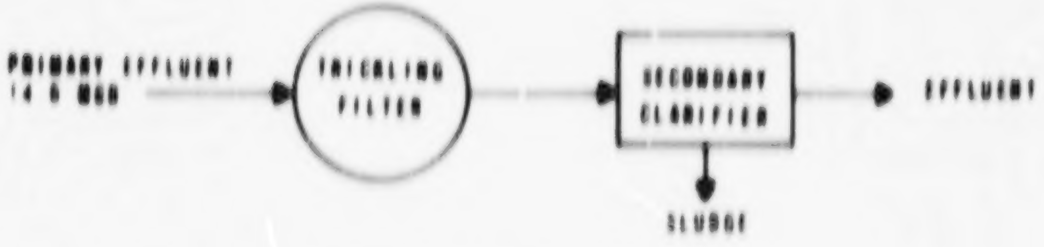
Recirculation pumps, motors and controls	$ 61,000
Piping and valves	68,000
Total	$129,000

4.5.1.2 Upgrading a Single-Stage Trickling Filter to a Two-Stage Filter System

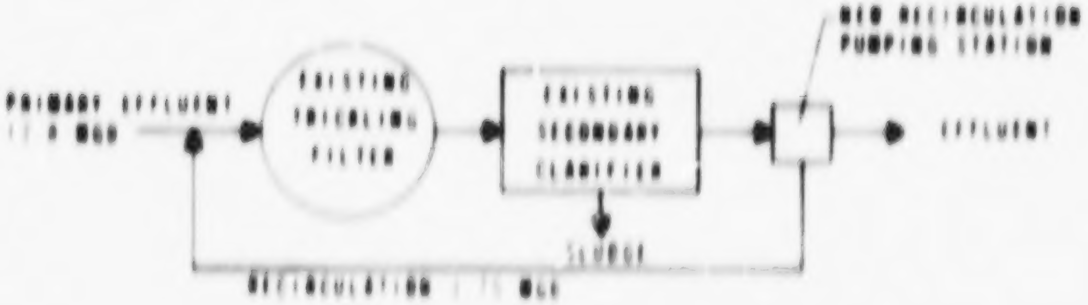
Upgrading an organically overloaded single-stage trickling filter may be accomplished by conversion to a two-stage filtration system. The objective of a pilot plant study conducted at the Chapel Hill, N.C., treatment plant in May-July, 1972, was to determine the degree of improvement that could be obtained by such a conversion (30).

FIGURE 4-3

UPGRADING A SINGLE-STAGE LOW-RATE TRICKLING FILTER BY IMPROVING DISTRIBUTION


PRIMARY EFFLUENT
14.8 MGD
TRICKLING
FILTER
SECONDARY
CLARIFIER
EFFLUENT
SLUDGE

TREATMENT SYSTEM BEFORE UPGRADING


PRIMARY EFFLUENT
[illegible] MGD
EXISTING
TRICKLING
FILTER
EXISTING
SECONDARY
CLARIFIER
NEW RECIRCULATION
PUMPING STATION
EFFLUENT
SLUDGE
RECIRCULATION [illegible] MGD

TREATMENT SYSTEM AFTER UPGRADING

TABLE 4-4
OPERATING DATA FOR PUEBLO, COLORADO

Description	Average Operating Condition (1959-66)	Upgraded Operating Condition (1973)
Flow, mgd	14.6	12.8
Influent BOD, mg/l	148	273
Influent SS, mg/l	163	122
Primary Clarifier		
Overflow Rate, gpd/sq ft	800	950
BOD Removal, percent	37	50
SS Removal, percent	49	49
Trickling Filter		
Depth, ft	6	6
Hydraulic Loading, mgd/acre[1]	9.4	10.7
Organic Loading, lb BOD/day/1,000 cu ft[1]	42	77
Recirculation Ratio	0	0.3
Secondary Clarifier		
Overflow Rate, gpd/sq ft[1]	775	800
BOD Removal, percent	50	74
SS Removal, percent	70	75
Overall Plant Performance		
BOD Removal, percent	74	86
SS Removal, percent	62	68
Effluent BOD, mg/l	39	37
Effluent SS, mg/l	59	40

[1] Includes recirculation

The single- versus two-stage studies were conducted using two pilot rock media trickling filter systems operated as shown on Figure 4-4. During the tests the influent flow to the single stage unit was maintained at 1.2 gpm while the two-stage influent flow rate was set at 2.4 gpm. These rates of flow were established so that performance could be compared for both a single-stage system where the plant influent is split into equal portions for treatment through the filters in parallel, and a two-stage system where the entire plant flow passes through the two filters in series. In the two-stage sequence, the entire influent flow was treated through one primary tank, then through one filter with recirculation to the head of primary. Effluent from the first-stage filter was then passed through the other primary clarifier, now serving as an intermediate clarifier. The partially treated flow was then fed to the second-stage filter with recirculation directly around the filter. Lastly, second-stage filter effluent was split evenly between the two final clarifiers. The operating data given in Table 4-5 show that 87 percent of the BOD and 93 percent of the SS were removed by the two-stage system. These were better reductions than the 80 percent BOD and 85 percent SS removals obtained with single-stage treatment.

4.5.1.3 Upgrading a Single-Stage Trickling Filter to a Two-Stage Biological Filtration/Activated Sludge System

If the hydraulic and organic loads to a high-rate trickling filter unit are such that it does not produce the desired effluent BOD quality, it is possible to upgrade the facility by the addition of an activated sludge unit immediately downstream from the existing filters. In this situation, the existing trickling filter acts as a roughing filter, and the subsequent activated sludge unit provides the treatment capacity needed to obtain the desired BOD removal.

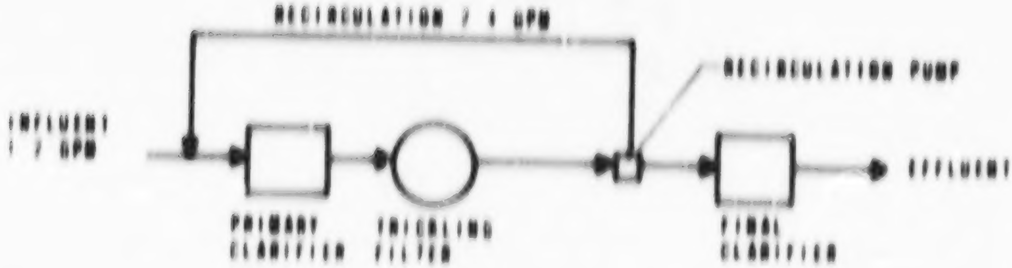
Alterations made on the Kankakee, Illinois, treatment plant beginning in 1969 exemplify this type of upgrading (15). The flow diagram of the plant before upgrading appears on Figure 4-5, and operating data for the year preceding the upgrading are summarized in Table 4-6. The plant was upgraded by conversion of the existing trickling filters to roughing filters, addition of a conventional activated sludge system with expanded sludge processing facilities, and modification as necessary to the existing piping. The flow diagram for the upgraded plant appears on Figure 4-5, and operating data for a year after the upgrading are given in Table 4-6. Implementation of this upgrading procedure improved the overall plant performance by increasing the BOD removal from 67 to 90 percent and the SS removal from 73 to 90 percent. Details concerning the design of an activated sludge system treating an effluent from a single-stage biological treatment process are presented in Chapter 5.

The construction costs for this plant modification were estimated to be $3,898,000 and were allocated as follows:

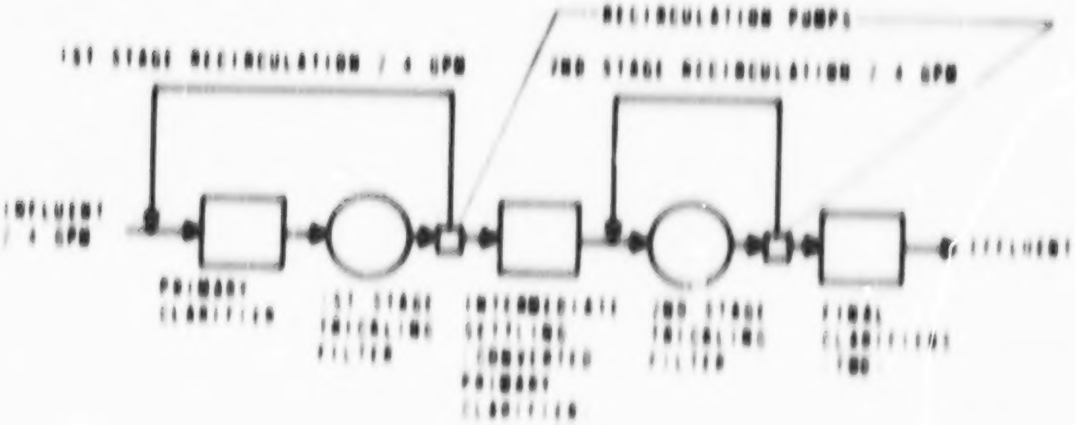
Materials and earthwork	$1,905,000
Piping and mechanical	814,000
Equipment	922,000
Electrical	257,000
Total	$3,898,000

FIGURE 4-4

MODIFYING A SINGLE-STAGE TRICKLING FILTER TO A TWO-STAGE FILTRATION SYSTEM


RECIRCULATION / 4 GPM
RECIRCULATION PUMP
INFLUENT
/ 7 GPM
EFFLUENT
PRIMARY
CLARIFIER
TRICKLING
FILTER
FINAL
CLARIFIER

TREATMENT SYSTEM PERFORMING AS
SINGLE STAGE TRICKLING FILTER SYSTEM
(ONE OF TWO DUPLICATE TRAINS)


RECIRCULATION PUMPS
1ST STAGE RECIRCULATION / 4 GPM
2ND STAGE RECIRCULATION / 4 GPM
INFLUENT
/ 4 GPM
EFFLUENT
PRIMARY
CLARIFIER
1ST STAGE
TRICKLING
FILTER
INTERMEDIATE
SETTLING
(CONVERTED
PRIMARY
CLARIFIER)
2ND STAGE
TRICKLING
FILTER
FINAL
CLARIFIERS
(2)

TREATMENT SYSTEM MODIFIED TO
TWO STAGE TRICKLING FILTER SYSTEM

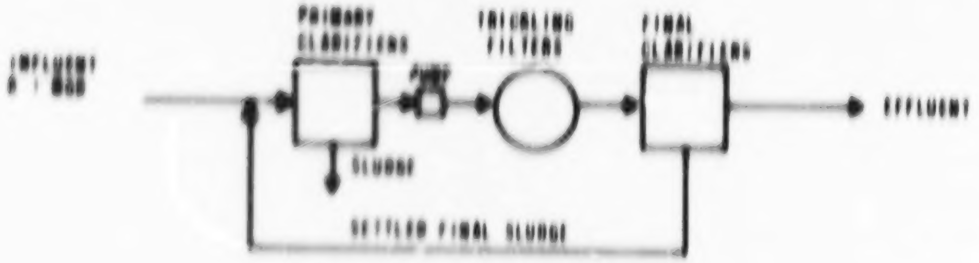
TABLE 4-5
OPERATING DATA FOR CHAPEL HILL, NORTH CAROLINA

Description	Single-Stage Operating Conditions	Two-Stage Operating Conditions
Flow, gpm	1.2	2.4
Influent BOD, mg/l	179	179
Influent SS, mg/l	247	247
Primary Clarifier		
Overflow Rate, gpd/sq ft	470	620
BOD Removal, percent	35	40
Trickling Filter - 1st Stage		
Depth, ft	4.25	4.25
Hydraulic Loading, mgd/acre[1]	18.0	23.9
Organic Loading, lb BOD/day/1,000 cu ft[1]	34.4	73.5
Recirculation Ratio	2.0	1.0
Final or Intermediate Clarifier		
Overflow Rate, gpd/sq ft[1]	436	314
BOD Removal - 1st Stage Filter and Clarifier, percent	69	59
Trickling Filter - 2nd Stage		
Depth, ft		4.25
Hydraulic Loading, mgd/acre[1]		23.9
Organic Loading, lb BOD/day/1,000 cu ft[1]		39.8
Recirculation Ratio		1.0
Final Clarifier		
Overflow Rate, gpd/sq ft		436
BOD Removal - 2nd Stage Filter and Clarifier, percent		55
SS Removal - 2nd Stage Filter and Clarifier, percent		44
Overall Plant Performance		
BOD Removal, percent	80	87
SS Removal, percent	85	93
Effluent BOD, mg/l	36	23
Effluent SS, mg/l	36	18

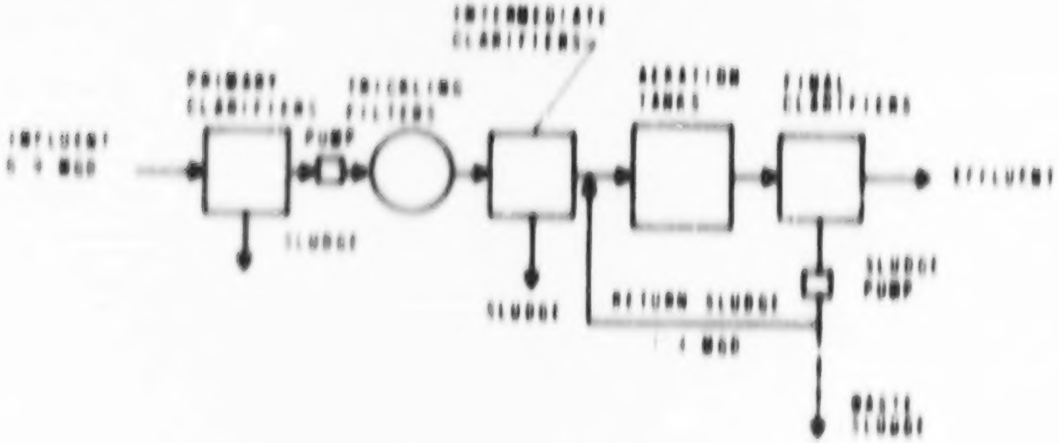
---

[1] Includes recirculation

FIGURE 4-5
UPGRADING A SINGLE STAGE TRICKLING FILTER TO A TWO-STAGE BIOLOGICAL FILTRATION/ACTIVATED SLUDGE SYSTEM


INFLUENT
[illegible] MGD
PRIMARY CLARIFIERS
PUMP
TRICKLING FILTERS
FINAL CLARIFIERS
EFFLUENT
SLUDGE
SETTLED FINAL SLUDGE

TREATMENT SYSTEM BEFORE UPGRADING


INTERMEDIATE CLARIFIERS
PRIMARY CLARIFIERS
TRICKLING FILTERS
AERATION TANKS
FINAL CLARIFIERS
INFLUENT
[illegible] MGD
PUMP
EFFLUENT
SLUDGE
SLUDGE
SLUDGE PUMP
RETURN SLUDGE
[illegible] MGD
WASTE SLUDGE

TREATMENT SYSTEM AFTER UPGRADING

TABLE 4-6
OPERATING DATA FOR KANKAKEE, ILLINOIS

Description	Year Before Upgrading	Year After Upgrading
Flow, mgd	6.1	6.9
Influent BOD, mg/l	292	313
Influent SS, mg/l	300	332
Primary Clarifier		
Overflow Rate, gpd/sq ft	700	600
BOD Removal, percent	21	47
SS Removal, percent	30	54
Trickling Filter		
Depth, ft	6.4	6.4
Hydraulic Loading, mgd/acre	13.5	11.3
Organic Loading, lb BOD/day/1,000 cu ft	93.4	56.9
BOD Removal, percent	28	37
Final (or Intermediate) Clarifier		
Overflow Rate, gpd/ft	1,030	900
BOD Removal - 1st Stage Filter and Clarifier, percent	43	80
SS Removal - 1st Stage Filter and Clarifier, percent	53	79
Complete Mixed Aeration Tank		
Detention Time Based on Average Flow, hr[1]	-	6.2
Sludge Recycle Rate, percent of Average Flow	-	30
Volumetric Loading, lb BOD/day/1,000 cu ft Aeration Tank Volume	-	25
Final Clarifier		
Overflow Rate, gpd/sq ft	-	610
BOD Removal - 2nd Stage Filter and Clarifier, percent	-	77
SS Removal - 2nd Stage Filter and Clarifier, percent	-	76
Overall Plant Performance		
BOD Removal, percent	67	98
SS Removal, percent	74	99
Effluent BOD, mg/l	95	6
Effluent SS, mg/l	75	4

[1] Includes sludge recycle.

4.5.1.4 Upgrading an Existing Two-Stage Trickling Filter to a Multiple Stage Biological System

Several options are available for upgrading a hydraulically or organically overloaded two-stage filter. Three of the more common techniques are listed below:

1. Construction of a roughing filter preceding the existing system
2. Construction of an activated sludge system following the existing system
3. Construction of a separate parallel biological treatment system

A detailed discussion will not be presented here, since most of the engineering considerations pertaining to these three options have been examined in previous sections.

4.5.2 Upgrading to Increase Organic Removal Efficiency

Upgrading techniques previously discussed relate to the ability of existing facilities to handle increased hydraulic or organic loads by providing modifications to meet existing effluent standards. However, there may be a need to meet higher effluent standards even though the existing facilities are not hydraulically or organically overloaded. Table 4.7 contains suggested alternatives for improving effluent quality under these conditions. The main purpose of the table is to present various alternatives and to suggest a range of anticipated improvement in performance for each alternative.

It should be emphasized that in cases where unit processes are added to existing facilities, the improvement in overall organic removal will be a direct function of the BOD removal achieved in the "add-on" unit process. However, where unit processes precede existing units, e.g., the use of a roughing filter, the overall BOD removal may not be increased in direct proportion to the amount achieved in the "add-on" process.

A detailed discussion on polishing lagoons, microscreens, filters, activated carbon and clarifier modifications appears in subsequent chapters. The applicability of alternatives to individual cases should be evaluated in detail prior to the implementation of a particular upgrading procedure.

4.5.2.1 Upgrading a Single Stage Trickling Filter Through Conversion to a Complete Mix Activated Sludge System

In 1965, the Ontario Water Resources Commission set 15 mg/l of BOD and SS as the objectives for secondary treatment plant effluents. This effluent quality could not be achieved with an existing high rate trickling filter plant at Gravenhurst, Ontario (12).

TABLE 4-7

UPGRADING TECHNIQUES FOR IMPROVEMENT OF TRICKLING FILTER PLANT EFFICIENCY

Addition Preceding Existing Unit	Modification to Existing Unit	Addition Following Existing Unit	Incremental BOD Removal Across the Added or Modified Process
			percent
	1. Low-Rate Trickling Filter		
	Add recirculation during low-flow periods		0-10
	2. High-Rate Trickling Filter		
	Increase recirculation		0-10
	3. Two-Stage Trickling Filter[1]		–
Roughing Trickling Filter (Rock or Synthetic Media)			20-40
Chemical Addition To Primary Clarifier			30-50
		2nd Stage Activated Sludge[2]	30-70
		Polishing Lagoon	30-60
		Multimedia Filter	50-80
		Microscreening	30-80
		Activated Carbon	60-80

[1] Generally not amenable to modifications for increasing treatment efficiency.

[2] A consideration if year-round nitrification is required.

To upgrade the high-rate filter, the plant was converted to complete mix activated sludge. The 60-foot diameter filter was converted to an aeration tank by removing the media and raising the concrete sidewalls 7 feet to a total height of 12 feet. A 40-hp mechanical aerator was installed. The dual clarifier (combination primary and secondary clarifier) was converted to a 60-foot diameter secondary clarifier, and a new 35-foot diameter primary clarifier was constructed. A 100-percent sludge recycle capacity was provided.

The previously described upgrading technique resulted in the following measured improvements:

Parameter	Before Upgrading	After Upgrading
Dry weather design flow, gpd	300,000	375,000
Influent organic load, lb BOD/day	360	540
Effluent BOD, mg/l	25-30	15-20

The upgrading technique employed allowed the plant to handle 50 percent greater BOD loads while producing an effluent of higher quality.

The capital costs for this upgrading were estimated at $82,000 and were allocated as follows:

Tank modification	$64,000
Secondary clarifier modification	18,000
Total	$82,000

These costs do not include upgrading of any other unit treatment processes, e.g., primary clarification.

4.5.3 Upgrading Existing Trickling Filters to Provide Nitrification

Increasingly stringent State and Federally approved water quality standards are requiring the partial or complete removal of nitrogenous oxygen demanding materials from critical stream basins, lakes and estuaries. For this reason, many existing trickling filter plants will have to be upgraded to provide nitrification.

The development and maintenance of nitrifying organisms in trickling filter systems is mainly dependent upon organic loading and wastewater temperature. Generally, nitrification occurs best at low BOD loadings and high wastewater temperature (20 deg C or higher). Table 4-8 clearly shows that the degree of nitrification in trickling filters improves as the volumetric BOD loading is reduced.

TABLE 4-5

TRICKLING FILTER NITRIFICATION DATA

Plant		Organic Loading BOD lb/day/1,000 cu ft	Influent BOD mg/l	Hydraulic Loading mgd/acre	Depth ft	Recirculation Ratio	Influent NH_3-N mg/l	Effluent NH_3-N mg/l	NH_3-N Removal percent	Reference
Livermore, Calif.		110	39	3.4	6.25	2.0	40.7	12.6	29	[illegible]
Glenwood City, Wisc.		90	168	14.3	7.0	2.0	10.6	8.7	17	36
Lakefield, Minn.		58	270	7.5	7.5	0.3	19.0	16.3	11	36
Allentown, Pa.		19	297	4.4	10.0	0.1	30.0	12.0	60	8
Ft. Benjamin Harrison, Ind.		6.0	113	1.7	8.0	0	21.0	6.3	70	[illegible]
Fitchburg, Mass.		3.7	180	6.0	10.0	0	12.0	2.0	83	38
Salford, England		3.2	290	0.7	8.0	0.1	30.0	6.7-0.4	91-98	37
		4.0	190	1.0	8.2	0.1	30.4	2.0-0.9	93-98	
		5.6	192	1.3	8.0	0.1	30.7	1.7-2.0	86-91	
		7.7	186	2.0	8.0	0.1	40.5	11.4-4.9	72-88	
		9.2	239	1.6	8.0	0.1	41.6	12.5-2.2	72-95	
		11.8	190	2.6	8.0	0.1	32.6	9.7-4.8	78-85	
		16.3	235	2.9	8.0	0.7	31.3	16.9-11.0	46-62	
		22.6	266	3.6	8.0	0.7	31.9	26.7-13.6	42-58	
Midland, Mich.	**Summer**	4.2-5.6	15-20	20	21.5	0-1.0	12.7	1.5	88	39
	Winter	4.2-5.6	15-20	40	21.5	1.0	14.0	2.2	84	

In colder climates, existing trickling filter plants may be modified to achieve a high degree of nitrification on a year-round basis or only during the warmer months of the year depending on effluent criteria. Year-round nitrification facilities must be designed for the lowest wastewater temperatures experienced in the winter months. In this instance, the required filter volume is generally much greater than that required for seasonal nitrification and normally requires at least two-stage treatment. Figure 4-6 illustrates four possible systems for implementing year-round nitrification.

Table 4-8 indicates that a high degree of nitrification (> 80 percent) was achieved at Salford, England (17) at BOD loading rates less than 12 and 6 lb BOD/day/1,000 cu ft with recirculation ratios of 1.0 and zero, respectively. These results were obtained with 20-foot diameter trickling filters, 6-feet deep, with a media consisting of blast furnace slag. At Midland, Michigan, a pilot study was conducted by Dow Chemical for the U.S. EPA (18), using synthetic media in a 3-foot diameter by 21.5-foot deep column. Table 4-9 shows that secondary effluent fed to the pilot facility at organic loading rates less than 50 lb BOD/day/1,000 cu ft produced a high degree of nitrification at temperatures as low as 7 deg C. The data shown in Table 4-8 and similar operating data from other locations indicate that trickling filters can achieve a high degree of nitrification at organic loadings less than 5 lb BOD/day/1,000 cu ft.

Example 1 on Figure 4-6 utilizes an activated sludge system which can also be used as the second-stage nitrification facility. This scheme can be used with or without an intermediate clarifier. If plant hydraulics permit, the existing first-stage clarifier may be converted for use as a final clarifier. The design of this type of facility is discussed in Chapter 5 and in reference (19).

The following examples describe the upgrading of existing trickling filter plants for nitrification.

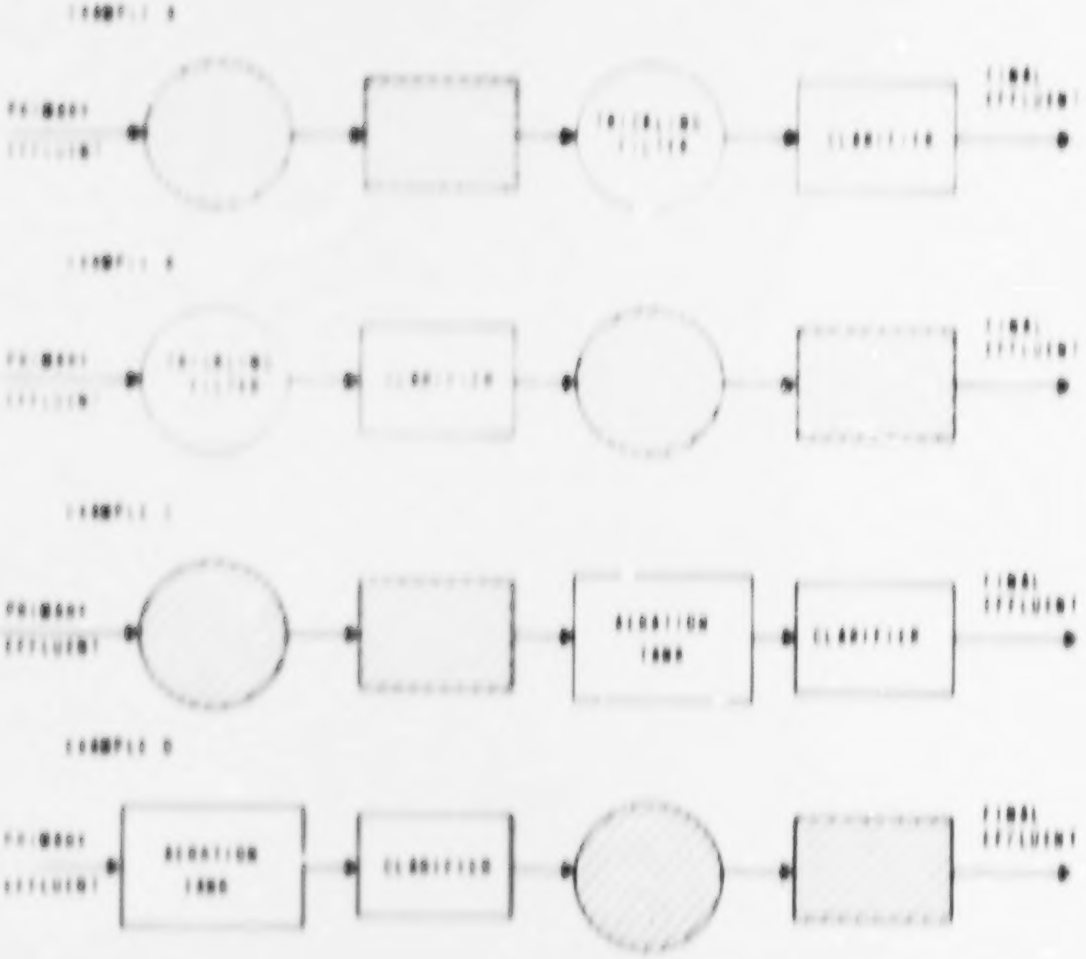
4.5.3.1 Upgrading a Single-Stage Trickling Filter to a Two-Stage Filtration System to Provide Nitrification

In January 1971, the Pennsylvania Department of Health upgraded the water quality criteria established for the Lehigh River and its major tributaries. As a result, the City of Allentown was informed that upgrading of its wastewater treatment plant would be required. The upgraded plant was designed to meet the following effluent standards from May 1 to October 31: BOD not to exceed 20 mg/l, ammonia nitrogen (NH_4-N) not to exceed 3 mg/l and total SS not to exceed 30 mg/l. It was also necessary to expand plant capacity from 20.5 to 30 mgd.

The existing plant provided secondary treatment using fixed-nozzle rock media trickling filters as shown on Figure 4-7. The raw wastewater to the plant contained 190 mg/l BOD, 16 mg/l NH_4-N, and 215 mg/l SS, which was reduced through the plant to effluent concentrations of 30 mg/l BOD, 12 mg/l NH_4-N, and 40 mg/l SS.

FIGURE 4.6

UPGRADING A TRICKLING FILTER SYSTEM TO PROVIDE TWO-STAGE NITRIFICATION

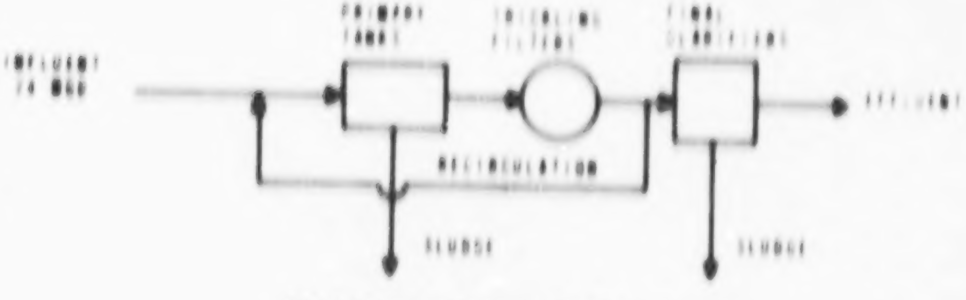

EXAMPLE A
PRIMARY EFFLUENT
TRICKLING FILTER
CLARIFIER
FINAL EFFLUENT
EXAMPLE B
PRIMARY EFFLUENT
TRICKLING FILTER
CLARIFIER
FINAL EFFLUENT
EXAMPLE C
PRIMARY EFFLUENT
AERATION TANK
CLARIFIER
FINAL EFFLUENT
EXAMPLE D
PRIMARY EFFLUENT
AERATION TANK
CLARIFIER
FINAL EFFLUENT

NOTES: 1. CROSS HATCHED FACILITIES DESIGNATE EXISTING TRICKLING FILTERS AND CLARIFIERS

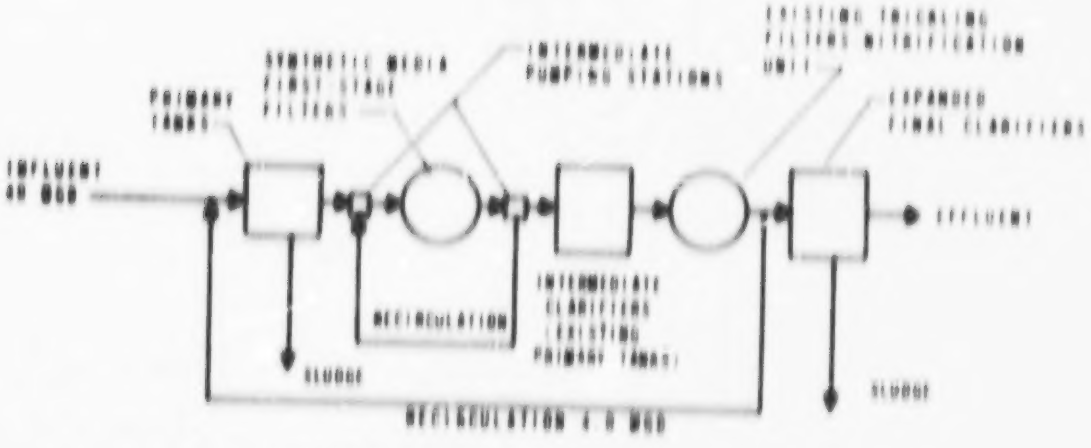
2. RECIRCULATION SCHEMES NOT SHOWN

FIGURE 4-7

UPGRADING A SINGLE STAGE TRICKLING FILTER TO A TWO STAGE FILTRATION SYSTEM TO PROVIDE NITRIFICATION


PRIMARY TANKS
TRICKLING FILTERS
FINAL CLARIFIERS
INFLUENT 14 MGD
EFFLUENT
RECIRCULATION
SLUDGE
SLUDGE

TREATMENT SYSTEM BEFORE UPGRADING


SYNTHETIC MEDIA FIRST STAGE FILTERS
INTERMEDIATE PUMPING STATIONS
EXISTING TRICKLING FILTERS NITRIFICATION UNIT
PRIMARY TANKS
EXPANDED FINAL CLARIFIERS
INFLUENT 40 MGD
EFFLUENT
INTERMEDIATE CLARIFIERS (EXISTING PRIMARY TANKS)
RECIRCULATION
SLUDGE
SLUDGE
RECIRCULATION 4.0 MGD

TREATMENT SYSTEM AFTER UPGRADING

Under the plan (33) recommended to meet the additional removal and flow requirements, an auxiliary pumping station was installed near the existing main station. The existing detritus tank and grit washing facilities were replaced by aerated grit chambers. New synthetic media first-stage filters were installed. The old rectangular primary tanks were used as intermediate clarifiers following the first-stage filters and new circular primary clarifiers were installed. The existing rock-media trickling filters were retained as nitrification units. The final clarifiers and chlorine contact tanks were expanded and modifications were made to the existing piping. Recirculation of flow to the trickling filters will only be used during low flow conditions to maintain a minimum hydraulic wetting rate. A flow diagram of the upgraded plant is shown on Figure 4-7. Operating conditions and upgraded design conditions are given in Table 4-9.

The capital costs associated with this upgrading were estimated at $11,642,000 and were allocated as follows:

Pumping	$ 1,192,000
Grit Handling Facilities	121,000
Primary Clarifiers	1,200,000
Trickling Filters	3,985,000
Final Clarifiers	115,000
Chlorine Contact Tank Additions	800,000
Sludge Processing	745,000
Instrumentation, Electrical, Plumbing and HVAC	1,647,000
Sitework, Outside Pumping and Dewatering	1,800,000
Miscellaneous	146,000
Total	$11,642,000

4.5.4.2 Upgrading a Single-Stage Trickling Filter to a Two-Stage System to Provide Nitrification

In 1969, an industry in New York State was required by the Department of Health to upgrade its existing trickling filter plant to meet year-round effluent standards of about 1600 lb/day of Ultimate Oxygen Demand (ultimate BOD plus nitrogenous oxygen demand). The plant consisted of single-stage, high-rate trickling filters designed to treat 0.9 mgd of sanitary and industrial wastes generated by the industry. The treatment system before upgrading is shown on Figure 4-8.

To meet the more stringent standard, it was decided to follow the existing facility with an activated sludge nitrification system (40). Except for some modifications in trickling filter recirculation, the existing facility was incorporated into the upgrading scheme unchanged. Effluent from the trickling filter plant was fed to a four-compartment, plug-flow nitrification reactor equipped with mechanical aerators. Two new final clarifiers followed the nitrification reactor. A flow diagram of the upgraded plant is shown on Figure 4-8.

TABLE 4-9
OPERATING AND DESIGN CONDITIONS FOR ALLENTOWN, PENNSYLVANIA

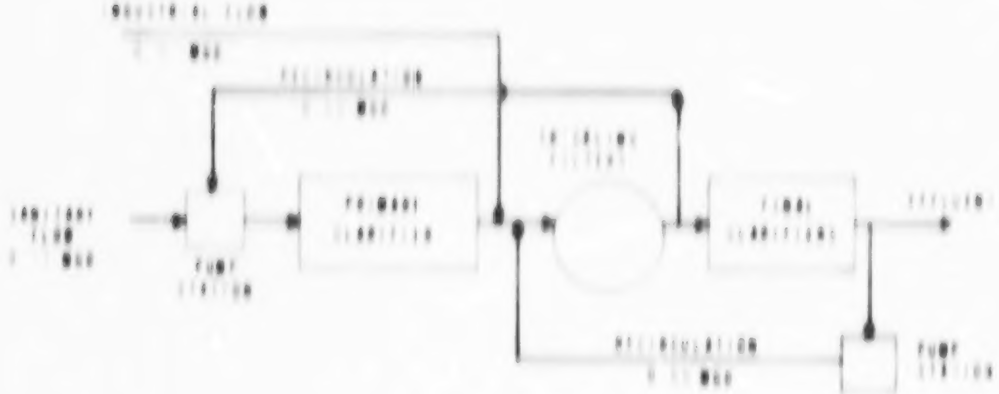
Description	1971 Operating Condition	Upgraded Design Condition
Flow, mgd	24	40
Influent BOD, mg/l	190	210
Influent SS, mg/l	215	230
Influent NH_3 - N, mg/l	16	15
Primary Clarifier		
Overflow Rate, gpd/sq ft	870	920
BOD Removal, percent		33
First Stage Filter - Synthetic Media		
Depth, ft	—	12
Hydraulic Loading, mgd/acre	—	58
Organic Loading, lb BOD/day/1,000 cu ft	—	50
Intermediate Clarifier		
Overflow Rate, gpd/sq ft	—	1,500
1st Stage BOD Removal, percent	—	60
1st Stage NH_3 - N Removal, percent	—	0
Trickling Filter - Rock Media[1]		
Depth, ft	10	10
Hydraulic Loading, mgd/acre	4.5	7.5
Organic Loading, lb BOD/day/1,000 cu ft	18.0	4.4
Final Clarifier		
Overflow Rate, gpd/sq ft	760	800
2nd Stage BOD Removal, percent	—	33
2nd Stage NH_3 - N Removal, percent	—	80
Overall Plant Performance		
BOD Removal, percent	79	91
SS Removal, percent	81	87
NH_3 - N Removal, percent	25	80
Effluent BOD, mg/l	40	20[2]
Effluent SS, mg/l	40	30[2]
Effluent NH_3 - N, mg/l	12	3[2]

[1] Proposed nitrification unit.

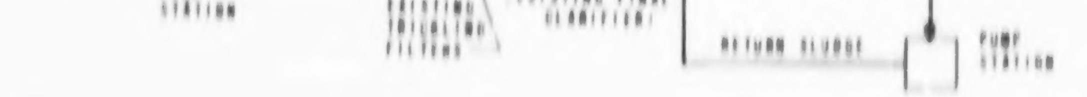
[2] May 1 - October 31.

FIGURE 4-8

UPGRADING A SINGLE STAGE TRICKLING FILTER TO A TWO-STAGE BIOLOGICAL SYSTEM TO PROVIDE NITRIFICATION


INDUSTRIAL FLOW
MGD
RECIRCULATION
MGD
TRICKLING FILTERS
SANITARY FLOW
MGD
PUMP STATION
PRIMARY CLARIFIER
FINAL CLARIFIERS
EFFLUENT
RECIRCULATION
MGD
PUMP STATION

TREATMENT SYSTEM BEFORE UPGRADING


INDUSTRIAL FLOW
2.64 MGD
RECIRCULATION
0.8 MGD
SANITARY FLOW
0.71 MGD
PUMP STATION
EXISTING PRIMARY CLARIFIER
EXISTING TRICKLING FILTERS
INTERMEDIATE CLARIFIER
(EXISTING FINAL CLARIFIER)
NITRIFICATION REACTOR
FINAL CLARIFIERS
EFFLUENT
RETURN SLUDGE
PUMP STATION

TREATMENT SYSTEM AFTER UPGRADING

Operating conditions and upgraded design conditions are given in Table 4-10. The upgraded facility has been in operation for over a year and has produced an effluent of higher quality than specified.

The capital cost for the construction of the upgraded facility was estimated at [illegible] in 1970.

4.5.4 Upgrading to Remove Nutrients

The preceding discussion has been primarily concerned with upgrading techniques to improve treatment plant effluent quality by reducing the oxygen demand that the effluent will exert on the receiving water. A matter of increasing concern, however, has been the nutrient content of treatment plant effluent. It is known that the continued and normal growth of microorganisms requires the availability of certain elements and nutrients as well as an energy source. Nitrogen and phosphorus are nutrients normally present in substantial quantities in raw wastewater that are not effectively removed by conventional biological treatment. The discharge of these substances contributes to the overfertilization and eutrophication of our surface waters. As a result, many states have established or are considering standards to limit the discharge of nutrients. The techniques that have been developed for nutrient removal are summarized in Table 4-11.

The design of a plant for nitrogen removal generally requires separate biological units for secondary treatment, nitrification and denitrification. In upgrading a trickling filter plant to remove nitrogen, the existing trickling filters may be used for roughing, as secondary treatment units, or as nitrification units.

The techniques available for removal of phosphorus are the subject of detailed discussion in the Process Design Manual for Phosphorus Removal (11). The following case histories illustrate the upgrading of existing trickling filter plants for phosphorus removal.

4.5.4.1 Upgrading a Low-Rate Trickling Filter System With Chemical Addition for Phosphorus Removal

A plant-scale study was begun at Richardson, Texas, in 1970 to evaluate the potential of chemical addition to remove phosphorus from and improve overall performance of the City's trickling filter plant (5). Major objectives were to reduce the effluent phosphorus concentration to 1.0 mg/l (as P), or less, and to reduce effluent BOD and SS residuals to 15 mg/l, or less.

The existing plant was a low-rate trickling filter facility with combination primary clarifier/digester units (clarigesters) as shown on Figure 4-9. Operating data for a control period before the addition of chemicals are shown in Table 4-12.

TABLE 4-10
OPERATIONAL AND DESIGN DATA FOR AN INDUSTRY IN NEW YORK STATE

Description	1970 Operating Condition	Upgraded Design Condition
Wastewater Flow, mgd		
Sanitary	0.11	0.25
Industrial	0.51	0.64
Total	0.62	0.89
Influent BOD, mg/l		
Sanitary	405	500
Industrial	55	120
Influent SS, mg/l		
Sanitary	224	150
Industrial	43	90
Primary Clarifier		
Overflow Rate, gpd/sq ft	336	435
BOD Removal (Sanitary Flow), percent	40	30
Trickling Filters - Stone Media		
Depth, ft	5	5
Hydraulic Loading, mgd/acre[1]	22.6	20.4
Organic Loading, lb BOD/day/1,000 cu ft[2]	34	86
Recirculation Ratio	1.67	0.67
Intermediate Clarifier (Existing Final Clarifier)		
Overflow Rate, gpd/sq ft[1]	740	560
1st Stage BOD Removal, percent	66	50
Nitrification Reactor		
Volume, cu ft		55,500
Depth, ft		12
Volumetric Loading, lb BOD/day/1,000 cu ft		12.5
MLSS, mg/l		4,000
F/M, lb BOD/day/lb MLSS		0.05
Sludge Recycle Rate, percent of Average Flow		100
Oxidizable Nitrogen to Reactor, mg/l		32.5

TABLE 4-10 (Continued)

Description	1970 Operating Condition	Upgraded Design Condition
Final Clarifier		
Overflow Rate, gpd/sq ft		800
2nd Stage BOD Removal, percent		92
Overall Plant Performance		
BOD Removal, percent	72	97
SS Removal, percent	44	>90
Effluent BOD, mg/l	36	7
Effluent SS, mg/l	42	<15
Effluent NH_4-N, mg/l	23	1.0

[1] Includes recirculation.

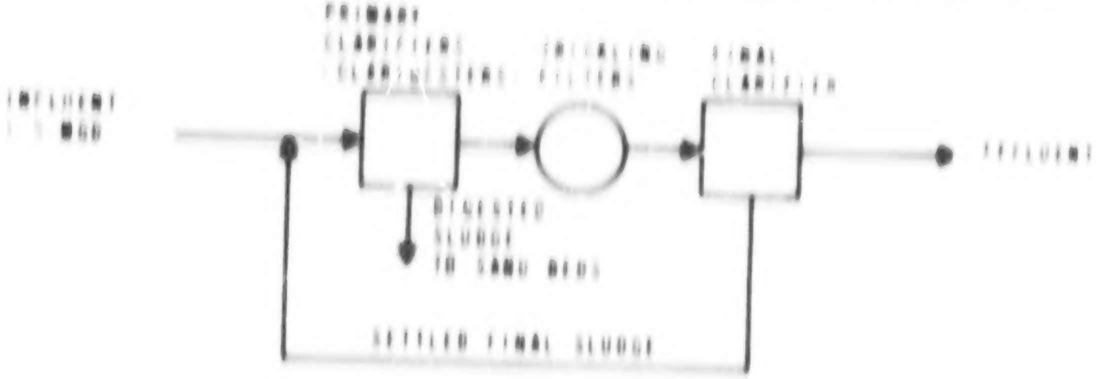
[2] Does not include recirculation.

TABLE 4-11
SUMMARY OF TREATMENT PROCESSES FOR NUTRIENT REMOVAL

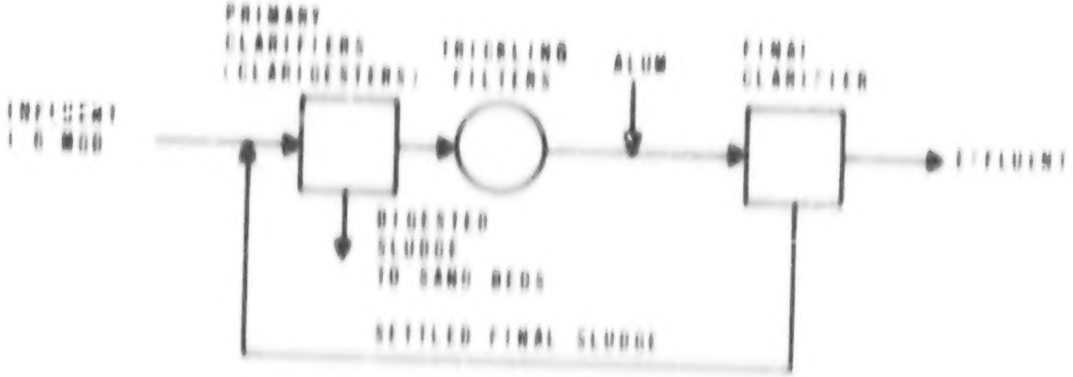
	Physical Chemical	Biological
Nitrogen	Ammonia Stripping after pH Adjustment	Nitrification-Denitrification
	Ammonia Removal by Breakpoint Chlorination	Algae Harvesting
	Ammonia Removal by Ion Exchange	
Phosphorus	Chemical Precipitation	

FIGURE 4.9

UPGRADING A STANDARD RATE TRICKLING FILTER SYSTEM WITH CHEMICAL ADDITION TO PROVIDE PHOSPHORUS REMOVAL


PRIMARY
CLARIFIERS
(CLARIGESTERS)
TRICKLING
FILTERS
FINAL
CLARIFIER
INFLUENT
1.0 MGD
EFFLUENT
DIGESTED
SLUDGE
TO SAND BEDS
SETTLED FINAL SLUDGE

TRICKLING FILTER SYSTEM BEFORE UPGRADING


PRIMARY
CLARIFIERS
(CLARIGESTERS)
TRICKLING
FILTERS
ALUM
FINAL
CLARIFIER
INFLUENT
1.0 MGD
EFFLUENT
DIGESTED
SLUDGE
TO SAND BEDS
SETTLED FINAL SLUDGE

TRICKLING FILTER SYSTEM AFTER UPGRADING

TABLE 4-12
OPERATING DATA FOR RICHARDSON, TEXAS

Description	1970 Control Period	1971-72 Extended Alum Run (11 1/2 Months)
Flow, mgd	1.5	1.6
Influent BOD, mg/l	166	170
Influent SS, mg/l	155	155
Influent P, mg/l	11	11.4
Primary Clarifier		
Overflow Rate, gpd/sq ft	800	425
Primary Effluent BOD, mg/l	*	115
Primary Effluent SS, mg/l	*	110
Primary Effluent P, mg/l	*	8.6
BOD Removal, percent	*	32
SS Removal, percent	*	29
P Removal, percent	*	25
Trickling Filter		
Depth, ft	6.5	6.5
Recirculation Ratio	0	0
Hydraulic Loading, mgd/acre	3.9	4.1
Organic Loading, lb BOD/day/1,000 cu ft	*	14.0
Secondary Clarifier		
Alum Dosage Rate, Average Al:P mole ratio	—	1.6**
Overflow Rate, gpd/sq ft	900	415
Secondary BOD Removal, percent	*	96
Secondary SS Removal, percent	*	94
Secondary P Removal, percent	*	94
Overall Plant Performance		
Effluent BOD, mg/l	20	5
Effluent SS, mg/l	15	7
Effluent P, mg/l	8	0.5
BOD Removal, percent	88	97
SS Removal, percent	90	95
P Removal, percent	27	96

*Not reported.

**Based on Influent P.

Liquid alum and liquid ferric chloride were selected as the two coagulants to be used in operational chemical addition trials. Points of chemical injection evaluated were just ahead of the influent wet well and in the junction box ahead of the final clarifier. The preliminary trials provided the following observations:

1. Ferric chloride addition at either injection point resulted in about 80 percent overall phosphorus removal, actually downgraded plant BOD and SS removals and produced excessive carry-over of discrete iron colloids in the plant effluent.

2. Alum addition to the influent wet well, although somewhat more effective than ferric chloride addition in achieving the desired effluent quality, resulted in a rapid and dramatic decrease in digester alkalinity and pH, necessitating termination of the trial in nine days.

3. Alum addition just ahead of the final clarifier proved to be the most effective chemical injection technique for removing phosphorus and upgrading BOD and SS removals, and did not exhibit the deleterious side effect on the unheated digesters observed when alum was dosed to the raw wastewater.

Based on the above operational trials, alum addition ahead of the final clarifier was selected as the most promising chemical additive system for an extended 11-1/2 month run. The flow diagram for the upgraded system is shown on Figure 4-9.

It was found that a mole ratio (Al : Influent P) of 1.5 to 1.7 consistently yielded effluent concentrations of 5 mg/l of BOD, 7 mg/l of SS and 0.5 mg/l of total phosphorus (as P). The corresponding effluent values for this plant prior to chemical addition were 30, 15 and 8 mg/l, respectively. The improved plant performance obtained with this upgrading technique was attributed in part to the low final clarifier overflow rate, careful management of final clarifier sludge withdrawal to prevent disruption to and loss of the alum floc blanket and frequent manual adjustment of the chemical feed pump rate to match alum dosage to mass inflow of phosphorus. Alum treatment doubled the volume of anaerobically digested sludge produced. However, the digested alum/biological sludge exhibited superior drying characteristics on sand beds and could be removed in about one-half the normal time. Operating data for the upgraded plant are shown in Table 4-12.

Chemical costs were $0.05/1,000 gal of plant flow or $0.36/lb of phosphorus removed, with phosphorus removal at the 96 percent level. The 1970 capital costs associated with plant modifications for chemical addition were $65,000, allocated as follows:

New laboratory building	$21,000
Laboratory equipment and furniture	6,000
Chemical storage and feed equipment, plant piping and metering modifications	38,000
Total	$65,000

4.5.4.2 Upgrading a High-Rate Trickling Filter System With Chemical Addition for *Phosphorus* Removal (A)

Promising results were obtained in removing phosphorus and in generally improving plant performance at Richardson, Texas, by the addition of alum to trickling filter effluent as discussed in Subsection 4.5.4.1. In view of these results, the University of North Carolina Wastewater Research Center initiated a follow-up study to further explore and confirm this process at the Chapel Hill, North Carolina Wastewater Treatment Plant (7). Conducting a similar chemical addition project at Chapel Hill was considered worthwhile because it is a typical high-rate trickling filter plant using recirculation, whereas the Richardson filters were low-rate units. Furthermore, parallel and identical lines of treatment units were available at Chapel Hill, allowing direct comparison of results with and without alum addition. Comparison of parallel results was not possible at Richardson.

The wastewater treatment plant at Chapel Hill is a conventional high-rate installation treating predominately domestic wastewater. Incoming wastewater passes through pretreatment facilities for the removal of large solids and grit. The flow is then divided into equal portions for diversion to two identical lines of treatment, each consisting of a primary clarifier, trickling filter and final clarifier as shown on Figure 4-10. On one side of the plant, a feed pump system was installed to add liquid alum to the trickling filter effluent just prior to the final clarifier.

Operating data for the two sides of the plant are shown in Table 4-13. Significant improvements in BOD, SS and phosphorus removals were achieved with an alum dosage rate that ranged between an Al:Influent P mole ratio of 1.5 and 2.2. Within this range and for the loading rates shown, the final clarifier hydraulic loading appeared to be the significant factor affecting process efficiency. Subsequent phases of the study indicated that phosphorus removals of over 90 percent and BOD removals of about 95 percent could be consistently obtained by lowering the final clarifier overflow rate to 500 gpd/sq ft. Recirculation of the final clarifier settled alum/humus sludge to the primary clarifier for thickening had a beneficial effect on primary treatment efficiency and decreased the organic loading on the alum train trickling filter.

Some problems were encountered at Chapel Hill in the sludge digestion and digester thickening operations during alum addition (see Section 11.2.6).

4.5.4.3 Upgrading a High-Rate Trickling Filter System With Chemical Addition for Phosphorus Removal (B)

In 1968, the City of Marlborough, Massachusetts, initiated a series of studies to improve the performance of its existing high-rate trickling filter plant (42). It was hoped that chemical treatment of the filter effluent prior to final settling would provide acceptable effluent concentrations of BOD, SS and phosphorus, with nominal additional capital costs. In 1971, following pilot studies, a plant scale program of alum addition was begun. A flow diagram of the existing single stage, high rate trickling filter system is shown on Figure 4-11.

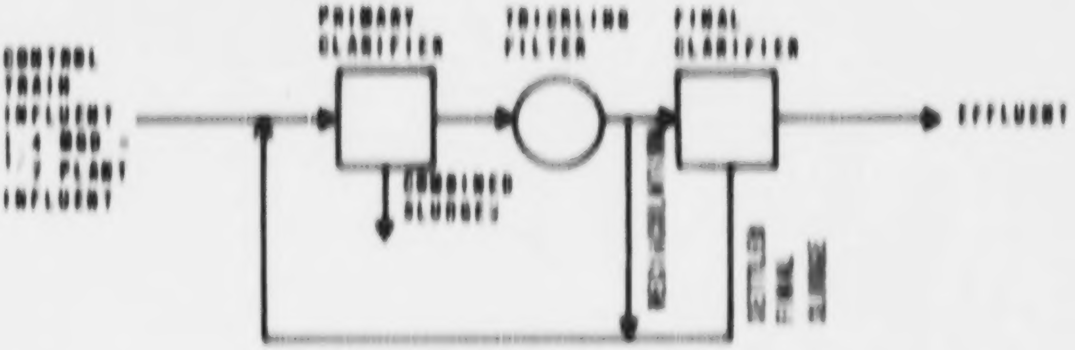
The existing plant includes Imhoff tanks for primary treatment and sludge digestion. Secondary humus sludge is returned to the Imhoff tanks' influent. Recirculation is accomplished with either trickling filter or final clarifier effluent. Pre-upgrading operating data for the period from January, 1970 to April, 1971 are shown in Table 4-14.

The upgrading modifications included the installation of alum storage and feeding equipment. Operating data for the upgraded plant from May, 1971 through August, 1972 are also shown in Table 4-14. As can be seen, the recycle of alum laden sludge from the final clarifier substantially improved the performance of the Imhoff tanks.

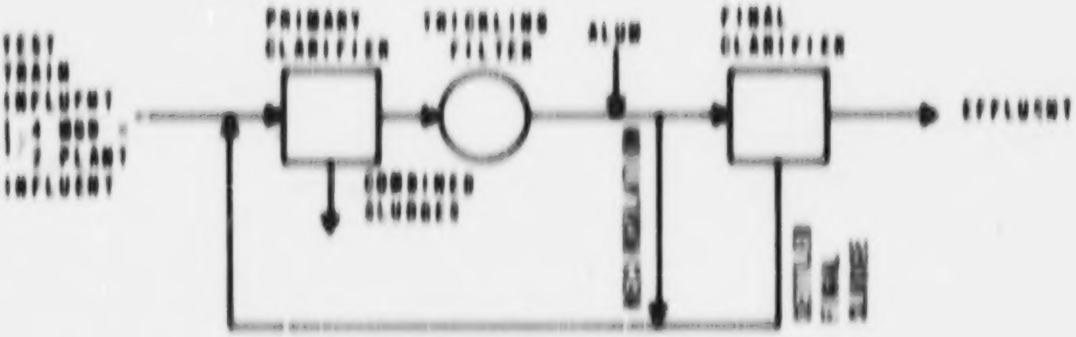
The capital cost for the chemical storage and feed system was $4,950. Based on a chemical cost of $65/ton of dry alum and a dosage of 1.5 moles of Al^{+3} to 1 mole of inorganic phosphorus, the treatment cost was $0.52/1,000 gal of plant flow.

FIGURE 4.10

UPGRADING A TRICKLING FILTER SYSTEM USING CHEMICAL ADDITION FOR PHOSPHORUS REMOVAL


CONTROL
TRAIN
INFLUENT
(.4 MGD =
1/2 PLANT
INFLUENT)
PRIMARY
CLARIFIER
TRICKLING
FILTER
FINAL
CLARIFIER
EFFLUENT
COMBINED
SLUDGES

TRICKLING FILTER TRAIN WITHOUT CHEMICAL ADDITION


TEST
TRAIN
INFLUENT
(.4 MGD =
1/2 PLANT
INFLUENT)
PRIMARY
CLARIFIER
TRICKLING
FILTER
ALUM
FINAL
CLARIFIER
EFFLUENT
COMBINED
SLUDGES

TRICKLING FILTER TRAIN WITH CHEMICAL ADDITION

TABLE 4-13
OPERATING DATA FOR CHAPEL HILL, NORTH CAROLINA

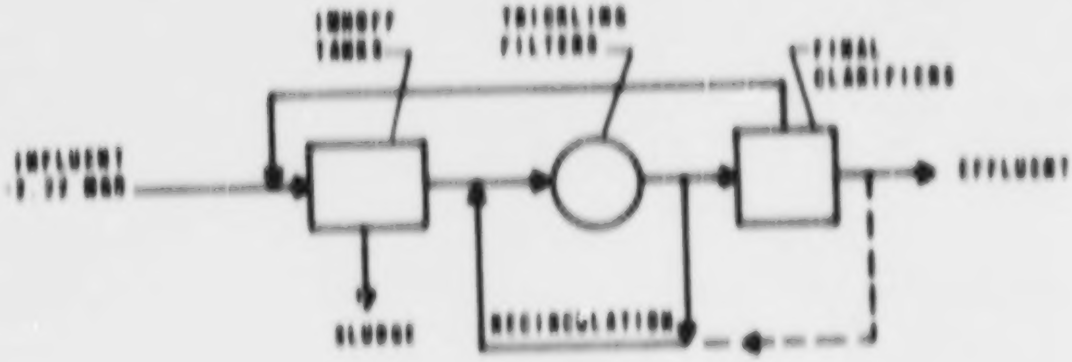
Description	Side 1 (No Alum)	Side 2 (With Alum)
Flow, mgd	1.41	1.41
Influent BOD, mg/l	168	168
Influent SS, mg/l	229	229
Influent P, mg/l	11.3	11.3
Primary Clarifier		
Overflow Rate, gpd/sq ft	1,100	1,100
Primary Effluent BOD, mg/l	77	62
Primary Effluent SS, mg/l	89	68
Primary Effluent P, mg/l	9.7	6.6
BOD Removal, percent	54	63
SS Removal, percent	61	70
P Removal, percent	14	42
Trickling Filter		
Depth, ft	4.25	4.25
Organic Loading, lb BOD/day/1,000 cu ft	2.0	2.0
Hydraulic Loading, mgd/acre[1]	16.3	16.3
Organic Loading, lb BOD/day/1,000 cu ft	18.6	15.4
Secondary Clarifier		
Alum Dosage Rate, Average Al:P mole ratio	-	1.7[2]
Overflow Rate, gpd/sq ft	885	885
Secondary BOD Removal, percent	43	76
Secondary SS Removal, percent	29	53
Secondary P Removal, percent	5	73
Overall Plant Performance		
Effluent BOD, mg/l	44	15
Effluent SS, mg/l	63	32
Effluent P, mg/l	9.2	1.8
BOD Removal, percent	74	91
SS Removal, percent	72	86
P Removal, percent	19	84

[1] Includes recirculation.

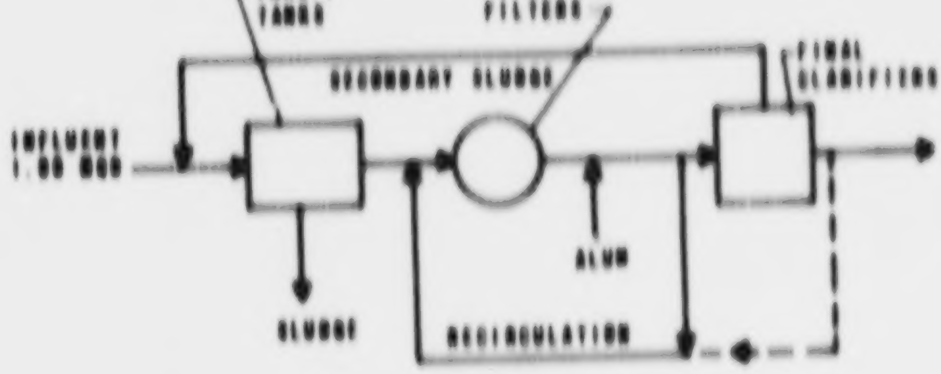
[2] Based on Influent P.

FIGURE 4-11

UPGRADING A TRICKLING FILTER SYSTEM TO PROVIDE PHOSPHORUS REMOVAL


IMHOFF
TANKS
TRICKLING
FILTERS
FINAL
CLARIFIERS
INFLUENT
0.77 MGD
EFFLUENT
SLUDGE
RECIRCULATION

TRICKLING FILTER SYSTEM BEFORE UPGRADING


IMHOFF
TANKS
TRICKLING
FILTERS
FINAL
CLARIFIERS
SECONDARY SLUDGE
INFLUENT
1.00 MGD
ALUM
SLUDGE
RECIRCULATION

TRICKLING FILTER SYSTEM AFTER UPGRADING

TABLE 4-14
OPERATING DATA FOR MARLBOROUGH, MASSACHUSETTS

Description	Operating Conditions Prior to Upgrading	Operating Conditions After Upgrading
Flow, mgd	2.28	1.99
Influent BOD, mg/l	134	171
Influent SS, mg/l	112	190
Influent Inorganic P, mg/l	7	8.1
Primary Settling - Imhoff Tanks		
Overflow Rate, gpd/sq ft	605	520
BOD Removal, percent[1]	3	22
Trickling Filter		
Depth, ft	6	6
Hydraulic Loading, mgd/acre[1]	21.4	19.1
Organic Loading, lb BOD/day/1,000 cu ft	74	87
Secondary Clarifier		
Alum Dosage Rate, Average Al : Inorganic P mole ratio	–	1.5
Overflow Rate, gpd/sq ft	915	785
BOD Removal, percent	54	77
Percent SS Removal percent[2]	71	78
Overall Plant Performance[3]		
BOD Removal, percent	70	86
SS Removal, percent	85	88
Inorganic P Removal, percent	14	81
Effluent BOD, mg/l	40	24
Effluent SS, mg/l	17	23
Effluent Inorganic P, mg/l	6	1.4

[1] Includes 0.8 mgd recirculation.

[2] Includes primary removal.

[3] Includes effluent polishing lagoon.

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CHAPTER 5

TECHNIQUES FOR UPGRADING ACTIVATED SLUDGE PLANTS

5.1 General

A better understanding of the activated sludge process has evolved down through the years, primarily due to an improved knowledge of the theory involved and to the experience accumulated in the successful operation of the process. For these reasons, it has become the most versatile biological treatment process available to the design engineer.

Historically, the activated sludge process has been used in larger cities, where the ratio of river assimilative capacity to waste load is small. More recently, there has been a trend toward its use by smaller communities to meet the more stringent requirements of regulatory agencies.

Existing overloaded conventional activated sludge plants pose a problem to the maintenance of established water quality standards. Various modifications of the conventional process developed over the years permit reduced aeration detention time and higher volumetric loadings. The applicability of these process modifications in the effort at upgrading of existing plants will be examined and discussed in this chapter.

5.2 Activated Sludge Processes

Basically, the activated sludge process uses microorganisms in suspension to oxidize soluble and colloidal organics to CO_2 and H_2O in the presence of molecular oxygen. During the oxidation process, a portion of the organic material is synthesized into new cells. A part of the synthesized cells then undergoes auto-oxidation in the aeration tank, the remainder forming excess sludge. Oxygen is required in the process to support the oxidation and synthesis reactions. To operate the process on a continuous basis, the solids generated must be separated in a clarifier with the major fraction being recycled to the aeration tank and the excess sludge being withdrawn from the clarifier underflow for further handling and disposal.

Due to the adaptability of the process, an activated sludge plant may be designed as a:

1. Conventional Plant
2. Step Aeration Plant
3. Contact Stabilization Plant
4. Complete Mix Plant
5. Modified Aeration Plant
6. Two-Stage Plant
7. Pure Oxygen Activated Sludge Plant.

It is possible to design a plant which can be operated under several of the above modifications.

When compared with one another, these plants each have advantages and disadvantages. Some achieve better BOD and SS removals than others, some cost less to construct, others cost less to operate, some produce less sludge and some obtain better nutrient removal. All of these factors must be considered in selecting a particular modification for use in a specific upgrading situation.

5.2.1 Conventional Activated Sludge

Design of a conventional activated sludge system is usually based upon volumetric loadings of 20 to 40 lb BOD/day/1,000 cu ft and organic loadings (food to microorganisms, F/M) of 0.2 to 0.4 lb BOD/day/lb MLSS. Sludge retention times (SRT) are normally between 5 and 15 days for this process which will achieve 85 to 95 percent BOD removals with proper operation (1). SRT as used herein is defined as:

$$\frac{\text{lb MLSS under aeration}}{(\text{lb SS wasted} + \text{lb SS lost in final effluent})/\text{day}}$$

A schematic for a conventional activated sludge plant is shown on Figure 5.1. The wastewater is commonly aerated for a period of 6 to 8 hours (based on the average design flow) in the presence of a portion of the secondary sludge (2). The rate of sludge return expressed as a percentage of the average wastewater design flow is normally about 25 percent, with minimum and maximum rates of 15 to 75 percent. A plug flow configuration is achieved in rectangular tanks, designed so that the total tank length is generally 5 to 50 times the width. Operating data from conventional activated sludge plants are summarized in Table 5.1.

The following factors have been recognized as limitations in the design and use of a conventional activated sludge plant:

1. Volumetric BOD loadings are limited to about 40 lb/day/1,000 cu ft because of poor load distribution.

2. Required aeration detention times are in the range of 6 to 8 hours.

3. A high initial oxygen demand is experienced in the head end of the aeration tank; however, this can be offset by tapering the air supply.

4. There may be a lack of operational stability with extreme variations in hydraulic and organic loadings.

TABLE 5-1

OPERATING DATA FROM CONVENTIONAL ACTIVATED SLUDGE PLANTS

[illegible]	[illegible]	[illegible]	[illegible]	[illegible]	[illegible]	[illegible]	[illegible]	[illegible]	[illegible]	[illegible]	[illegible]
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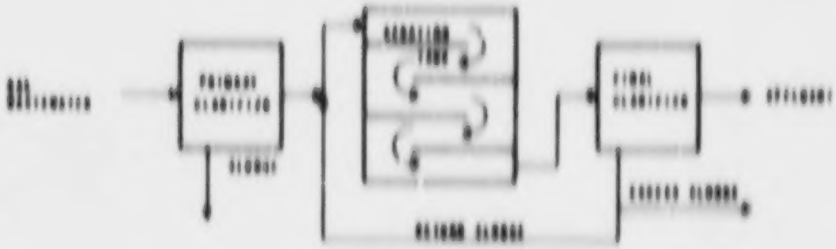
FIGURE 5-1

CONVENTIONAL ACTIVATED SLUDGE PLANT



Some of these limitations have stimulated the development and use of various process modifications, such as step aeration, contact stabilization, complete mix, modified aeration, two-stage activated sludge and the use of oxygen instead of air as a source of DO. These process modifications are discussed in subsequent sections.

5.2.2 Step Aeration

A typical flow diagram of a step aeration plant is illustrated on Figure 5.2. Unlike the conventional plant, the influent wastewater is introduced at several points along the aeration tank. However, the return sludge, which normally ranges from 25 to 75 percent of the average design flow, is introduced at the head end of the aeration tank as in the conventional system. Distributing the influent flow along the aeration tank reduces the initial oxygen demand usually experienced in the conventional plant. This permits a more efficient utilization of the activated sludge biomass.

Step aeration plants are usually designed for volumetric loadings of 40 to 60 lb BOD/day/1,000 cu ft at F/M's varying from 0.2 to 0.4 lb BOD/day/lb MLSS. SRT's are similar to those of the conventional system. A step aeration system will achieve 85 to 95 percent BOD removal. Operating data from step aeration plants are summarized in Table 5.2.

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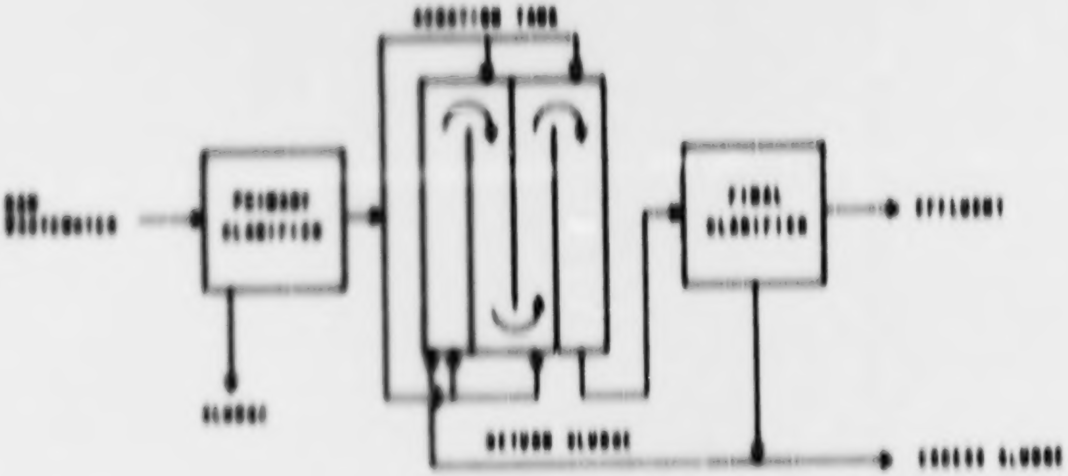
FIGURE 5-2

STEP AERATION PLANT



Biological oxygen requirements for step aeration are similar to those of conventional activated sludge. However, because of more uniform load distribution, the actual air supplied is more effectively utilized and the quantity may be somewhat reduced. Since the detention times are lower than for the conventional system, the air piping and diffusion equipment must be modified to supply approximately the conventional volume of air to a tank approximately one-half the conventional size.

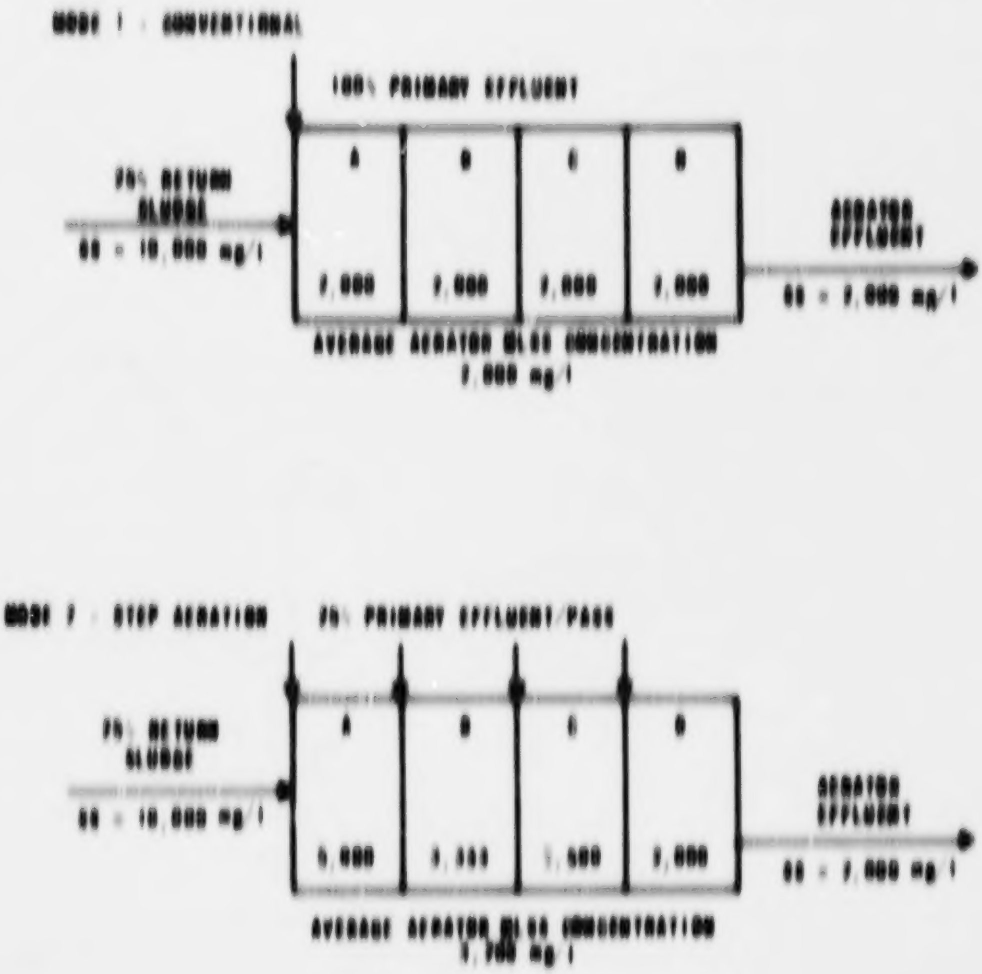
In the conventional plant, the MLSS concentration is relatively constant throughout the aeration tank; in the step aeration plant, the MLSS concentration decreases at each point of influent addition. The reason that a step aeration system can be operated at the same F/M as a conventional system, but in about one-half to two-thirds of the aeration tankage, is the higher average MLSS concentration gained by the incremental addition of flow. The step feed concept permits the utilization of the same size clarifiers as the conventional system because the aerator effluent SS concentrations are similar. This principle is shown on Figure 5-3. The concept can also be used to decrease the final clarifier solids loading while maintaining the same average aerator MLSS concentration.

5.2.3 Contact Stabilization

The principles of the contact stabilization modification were initially demonstrated in the upgrading of an existing hydraulically overloaded conventional plant (10). As in the step

FIGURE 5.3

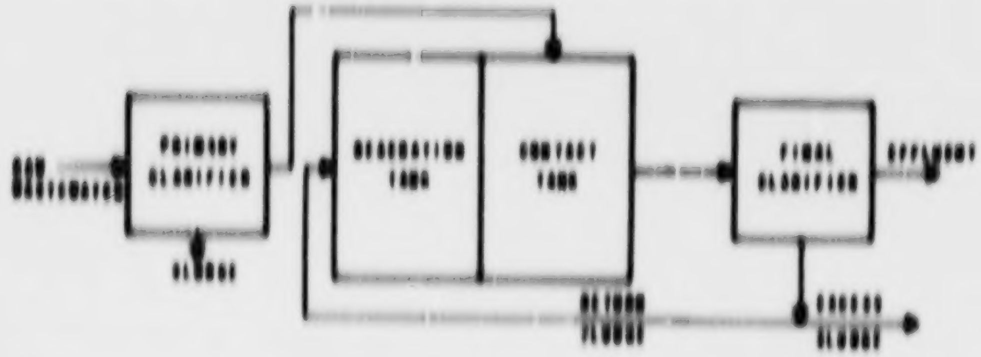
COMPARISON OF SOLIDS LOADING ON THE FINAL CLARIFIER FOR CONVENTIONAL AND STEP AERATION PLANTS


MODE 1 - CONVENTIONAL
100% PRIMARY EFFLUENT
25% RETURN SLUDGE
SS = 10,000 mg/l
A
B
C
D
2,000
2,000
2,000
2,000
AERATED EFFLUENT
SS = 2,000 mg/l
AVERAGE AERATOR MLSS CONCENTRATION
2,000 mg/l
MODE 2 - STEP AERATION
25% PRIMARY EFFLUENT/PASS
25% RETURN SLUDGE
SS = 10,000 mg/l
A
B
C
D
5,000
3,333
2,500
2,000
AERATED EFFLUENT
SS = 2,000 mg/l
AVERAGE AERATOR MLSS CONCENTRATION
3,200 mg/l

NOTE: EXAMPLE ASSUMES NEGLIGIBLE SS IN PRIMARY EFFLUENT aeration process, this modification involves a change in the feed location to the aeration tanks. Volumetric BOD loadings, F/M, SRT and removal efficiency are similar to those of the step aeration system. Sludge return ratios vary from 25 to 100 percent of the average design flow. A schematic for the contact stabilization process is shown on Figure 5-4.

FIGURE 5-4

CONTACT STABILIZATION PLANT



Laboratory studies and field work have demonstrated that wastewater BOD in the colloidal or insoluble state is rapidly removed from wastewater in a relatively short contact time by the combined mechanism of biological sorption, synthesis and flocculation. This may offer the possibility of a reduction in plant volume for wastewaters exhibiting these characteristics. In the contact stabilization process, after the biological sludge is separated from the wastewater in the clarifier, the concentrated sludge is separately aerated in a reaeration tank. Here the flocculated and absorbed BOD is stabilized. In addition to a smaller total aeration volume than required with the conventional activated sludge process, the contact stabilization process has the advantage of being able to handle greater shock and toxic loadings because of the biological buffering capacity of the reaeration tank, and the fact that at any given time the majority of the activated sludge is isolated from the main stream of the plant flow. Operating data from contact stabilization plants are summarized in Table 5-6.

The detention times required in the sludge reaeration and contact tanks are interdependent. The contact tank detention time also depends on wastewater characteristics. For domestic wastewaters containing normal amounts of insoluble and colloidal BOD, contact tank

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[illegible]

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5. [illegible]

detention times of about 0.5 to 1 hour are employed based on average plant flow, with reaeration times of 3 to 6 hours based on sludge recycle flow. Total air requirements for this process are similar to those of conventional activated sludge and are normally divided equally between the contact and reaeration tanks.

Most of the benefits of contact stabilization are achieved if the organic load is present mainly in the colloidal state. Generally, the greater the fraction of soluble BOD, the greater the required contact time. As a result, the required total aeration volume of this process approaches that of the conventional process as the relative amount of soluble BOD in the wastewater increases.

Ten-States Standards require significantly higher contact and reaeration times than those previously cited, especially for smaller sized plants as indicated in Table 5-4 (14). These standards also specify an F/M of 0.2 to 0.5 lb BOD/day/lb MLSS and a volumetric loading of 30 to 50 lb BOD/da./1,000 cu ft. Although these values were, no doubt, selected to compensate for the extreme flow variations that occur at small plants, their use may result in poor quality effluents (15). McKinney (16) has indicated that in typical conservatively designed contact stabilization plants, all of the stabilization of the organic matter in the raw wastewater occurs in the contact zone; therefore, only endogenous respiration occurs in the reaeration tank. This situation often results in poor sludge settling characteristics in the secondary clarifier.

TABLE 5-4
SUGGESTED DESIGN GUIDELINES (14)

Plant Design Flow mgd	Contact Time[1] hours	Reaeration Time[2] hours	Aerator Loading[3] lb BOD/day 1,000 cu ft
to 0.5	3.0	6.0	30
0.5 to 1.5	3.0 to 2.0	6.0 to 4.0	30 to 50
1.5 and up	2.0 to 1.5	4.0 to 3.0	50

[1] Based on average design flow.
[2] Based on average sludge recycle flow.
[3] Based on total aeration capacity (reaeration plus contact).

5.2.4 Complete Mix

In the complete mix process, influent wastewater and recycled sludge are introduced uniformly throughout the aeration tank, as indicated on Figure 5.5. This flow distribution results in a uniform oxygen demand throughout the aeration tank which adds some operational stability when treating slug loads of industrial wastes (17). This process may be loaded to levels comparable to those of the step aeration and contact stabilization processes with only slight reductions from the removal efficiencies of these processes. The reduced efficiency occurs because there is a small amount of short circuiting in the complete mix aeration tank. Operating data from plants utilizing complete mix are presented in Table 5.5.

5.2.5 Modified Aeration

The flow scheme for modified aeration is similar to that of conventional activated sludge (Figure 5.1) with the exception that this process is frequently used without primary settling. In this system, organisms are brought into contact with the incoming wastes for a brief period of sorption and synthesis. Modified aeration plants are designed to process volumetric BOD loadings varying from 75 to 150 lb BOD/day/1,000 cu ft, and will achieve 60 to 75 percent BOD removals under these loading conditions. F/M's normally range from 1.5 to 5.0 lb BOD/day/lb MLSS, and SRT's vary from 0.2 to 0.5 day. For proper operation, the return sludge rate normally ranges from 5 to 15 percent of the influent flow (1) (24). Aeration requirements are less than those of the conventional system due to the reduced biological activity.

Modified aeration has been proposed as the first stage in a two-stage aeration system where the second stage will be used for nitrification. Two advantages of using modified aeration in the first stage are a substantial cost savings in aeration facilities and an SRT sufficiently low to preclude nitrification. The absence of nitrification and the partial concentration damping afforded by mixing in the modified aeration tank results in a more uniform ammonia concentration in the influent to the nitrification system (1) (25).

When modified aeration is used as the first stage of a two-stage system, the first-stage effluent BOD should be sufficiently high to permit good sludge flocculation and settling in the second-stage clarifier, but not so high as to cause a high heterotroph growth rate and a subsequent washout of the nitrifying population in the second stage. The upper limit will depend on the BOD/NH_3 of the second-stage influent. At the U. S. EPA District of Columbia Pilot Plant, addition of alum helped to maintain the BOD of the first-stage effluent within the desired range of 50-60 mg/l (25). Operation of the full scale District of Columbia Wastewater Treatment Plant has also shown that oxygen requirements for the modified process are significantly less than for other activated sludge systems and that the sludge exhibits good settling qualities (26) (25).

Operating data from modified aeration plants are shown in Table 5.6.

FIGURE 5.5

COMPLETE MIX PLANT

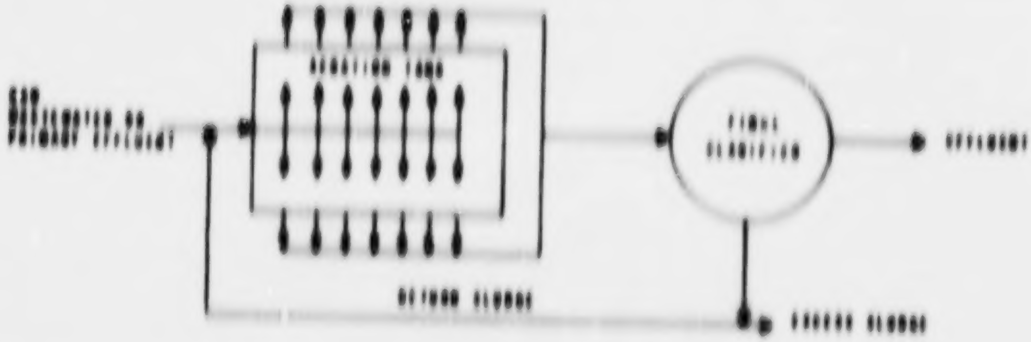
RAW
WASTEWATER OR
PRIMARY EFFLUENT
AERATION TANK
FINAL
CLARIFIER
EFFLUENT
RETURN SLUDGE
EXCESS SLUDGE

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TABLE 54

OPERATING DATA FROM MODIFIED AERATION ACTIVATED SLUDGE PLANTS

[illegible]

5.2.6 Two-Stage Activated Sludge

A two-stage activated sludge plant is essentially two separate activated sludge processes operating in series, as shown on Figure 5.6. The two separate sludge systems permit the development of two specialized microbial populations. In the first stage, the bulk of the carbonaceous material is removed by a wide variety of heterotrophic organisms commonly found in activated sludge. The reduction of BOD in the first stage permits an accumulation of the slower growing nitrifying organisms in the second stage which oxidize the ammonia nitrogen to the nitrate form.

The operating data presented in Table 5.7 indicate that the second-stage effluent BOD is not sufficiently better than for alternative single stage systems previously discussed to warrant consideration of this approach unless nitrification is also required. The advantage of satisfying the oxygen demand of the ammonia nitrogen normally discharged should not be underestimated since this can amount to as much as 70 percent of the total oxygen demand of the plant secondary effluent (30). The high air requirement in lb air supplied/lb BOD removed in the second stage includes the air used for ammonia oxidation.

5.2.7 Pure Oxygen Activated Sludge

The use of pure oxygen for activated sludge treatment has become competitive with the use of air due to the development of efficient oxygen dissolution systems. Efficient utilization of oxygen can be achieved by two methods:

1. Oxygenation is performed in a staged, covered reactor in which oxygen gas is recirculated within the system until it reaches a level of impurity at which it can no longer be used. This method is presently used in several municipal and industrial treatment plants, and is shown schematically on Figure 5.7 (31).

2. Oxygenation is performed in an open reactor in which extremely fine diffusers are utilized to develop small oxygen gas bubbles that are completely dissolved before breaking surface in normal-depth tanks (31). This approach has not yet been implemented in full scale treatment plants.

The use of pure oxygen has the advantages of reduced reactor volume, high effluent DO and effective odor control. Possible additional advantages include less waste sludge production and higher waste sludge concentrations. Disadvantages include the increased complexity of operation of both oxygen generation and dissolution systems and excessive pH depression in low alkalinity wastewaters when alum is added for phosphorus removal or when nitrification occurs.

Oxygen gas can be produced at the plant site by either a cryogenic unit or, in the case of smaller plants, a molecular sieve device. A liquid oxygen storage tank is generally

FIGURE 5-6
TWO-STAGE ACTIVATED SLUDGE PLANT

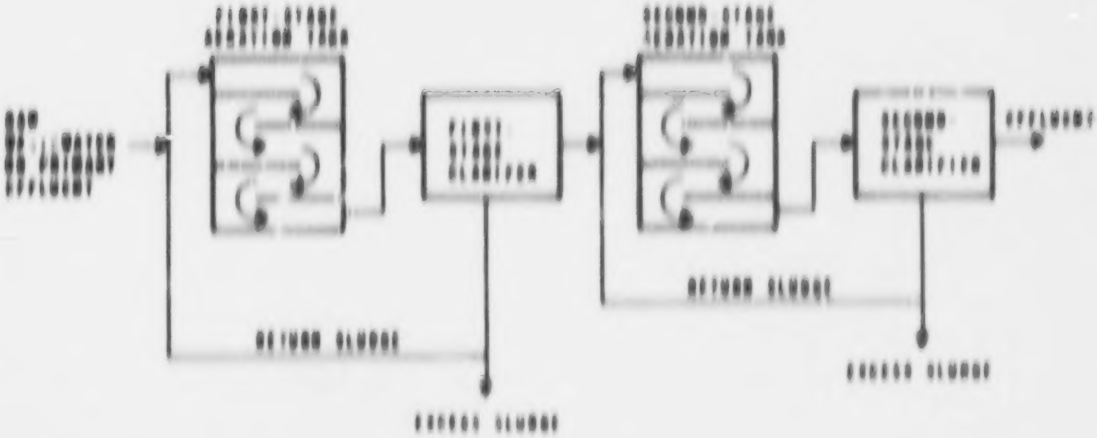

FIRST STAGE AERATION TANK
FIRST STAGE CLARIFIER
SECOND STAGE AERATION TANK
SECOND STAGE CLARIFIER
EFFLUENT
RETURN SLUDGE
RETURN SLUDGE
EXCESS SLUDGE
EXCESS SLUDGE

TABLE 5-7

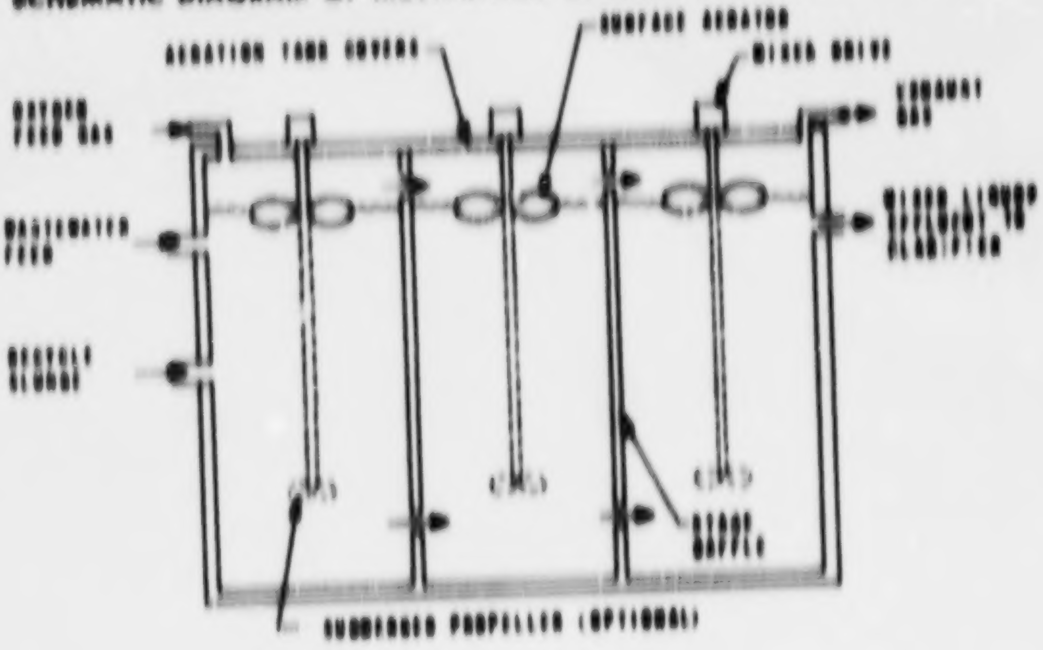
OPERATING DATA FROM TWO-STAGE ACTIVATED SLUDGE PLANTS

1st STAGE PERFORMANCE

2nd STAGE PERFORMANCE

FIGURE 5-7

SCHEMATIC DIAGRAM OF MULTI-STAGE OXYGEN AERATION SYSTEM (81)


SURFACE AERATOR
AERATION TANK COVERS
MIXER DRIVE
OXYGEN FEED GAS
EXHAUST GAS
WASTEWATER FEED
MIXED LIQUOR EFFLUENT TO CLARIFIER
RECYCLE SLUDGE
STAGE BAFFLE
SUBMERGED PROPELLER (OPTIONAL)

recommended to provide standby and peak load capacity. In the covered reactor design, oxygen transfer and mixing are accomplished with surface aerators for tank depths up to 15 to 18 feet (as shown on Figure 5.7) or alternatively with submerged turbine spargers and recirculating gas compressors for deeper tanks.

Pure oxygen systems are normally designed to handle volumetric BOD loadings ranging from 100 to 225 lb BOD/day/1,000 cu ft, but in some instances much higher loadings have been reported (51)(52)(53)(54)(55)(56).

The F/M may vary from 0.07 to more than 1.0, but normal design values range from 0.5 to 0.8 (51)(52)(53). Table 5.6 summarizes operating data from plants utilizing pure oxygen.

Pure oxygen pilot plant studies have indicated that at a given F/M oxygen consumption increases with higher COD/BOD ratios, as shown on Figure 5.8 (53). This increase in oxygen requirements may be attributed to the biodegradation of part of the COD which is not measured in the standard BOD test. Oxygen requirements for pure oxygen systems for typical domestic wastewaters normally range from 0.6 to 0.7 lb O_2/lb COD removed in the system (51).

Since the pure oxygen system operates at high MLSS concentrations, the clarifier will often be subjected to increased solids loadings. Clarifier solids loadings for pure oxygen and other systems are discussed in detail in Chapter 6. Clarifier underflow SS concentrations may range from 1.0 to 2.5 percent, depending on the sludge composition and the clarifier sludge detention time.

5.8 Activated Sludge Design Considerations

Initially, plant operators, through a trial and error procedure, developed the most efficient operating criteria for conventional activated sludge plants as well as for the modifications of the process. Out of this evolution, basic design criteria were developed. These criteria are still in use today and, in many cases, are rigidly adhered to by regulatory agencies.

A limitation of these design criteria is that volumetric loading (lb BOD applied/day/1,000 cu ft) has been considered preferable, for design purposes, to organic or F/M loading (lb BOD applied/day/lb MLSS). Many of the modifications have shown that organic loading is an important consideration; in fact, a higher volumetric loading has been achieved for modifications of the conventional process at the same organic loading used in the conventional process.

Basic parameters of interest in the design of an activated sludge process are:

1. BOD removal rates
2. Oxygen and air requirements

[illegible]

[illegible]

[illegible]	[illegible]	[illegible]	[illegible]		[illegible]	[illegible]	[illegible]	[illegible]	[illegible]	[illegible]
			[illegible]	[illegible]						
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	[illegible]	[illegible]	[illegible]	[illegible]	[illegible]	[illegible]	[illegible]	[illegible]	[illegible]	
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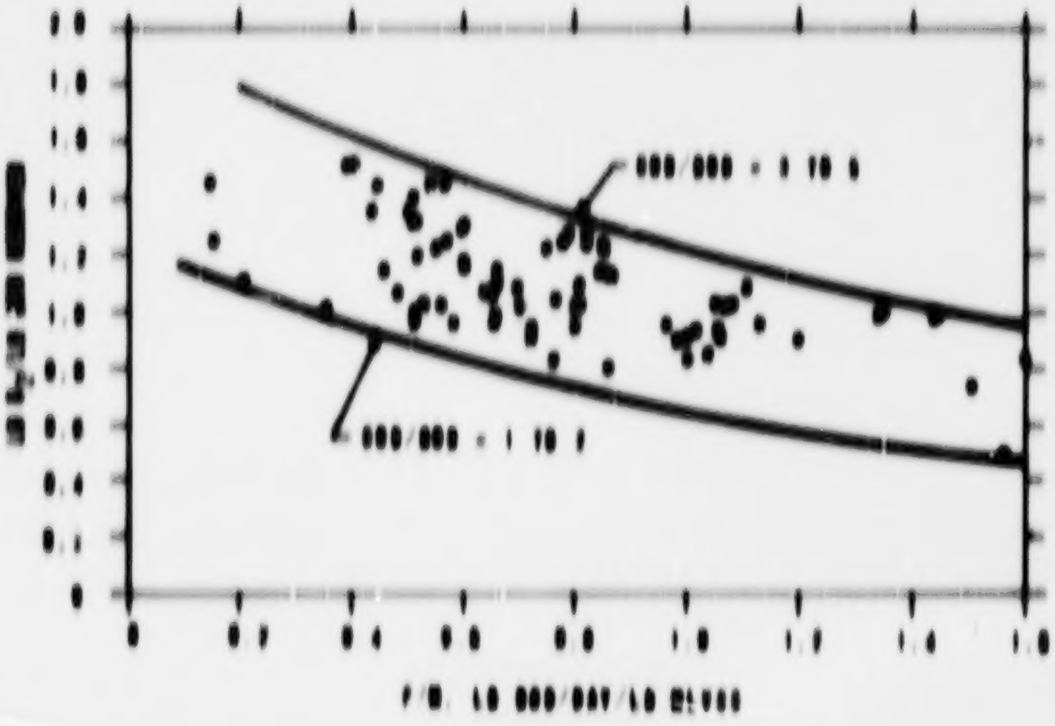
[1] [illegible]

[2] [illegible]

[3] [illegible]

FIGURE 5-6

OXYGEN CONSUMPTION FOR PURE OXYGEN SYSTEMS RELATED TO ORGANIC LOADING (F/M) (58)



3. Sludge production
4. Oxygen transfer rates in wastewater
5. Nutrient requirements
6. Separation and return of activated sludge.

5.5.1 BOD Removal Rates

Eckenfelder (41) has indicated that a linear arithmetic relationship exists between BOD removal rate (mg BOD/hr/g VSS) and effluent BOD (mg/l), as shown on Figure 5-9 for typical data from various complete-mix activated sludge plants. The variations in the BOD removal relationship on Figure 5-9 are influenced by the presence of various proportions of domestic and industrial wastes.

Using Figure 5-9, the detention time required to achieve a specific effluent BOD can be obtained as follows:

$$t = \frac{L_a - L_e}{S_a \, r'}$$

where:

t = Aeration tank detention time, hours
L_a = Influent BOD to aeration tank, mg/l
L_e = Clarifier effluent BOD, mg/l
S_a = MLVSS, mg/l
r' = BOD removal rate, mg BOD/hr/g VSS

Weston (42) developed a log-log relationship between a BOD removal rate constant (r) and a loading ratio (L_m/S_a); r and L_m are defined by the following equations:

$$r = \frac{L_m - L_e}{t_a S_a}$$

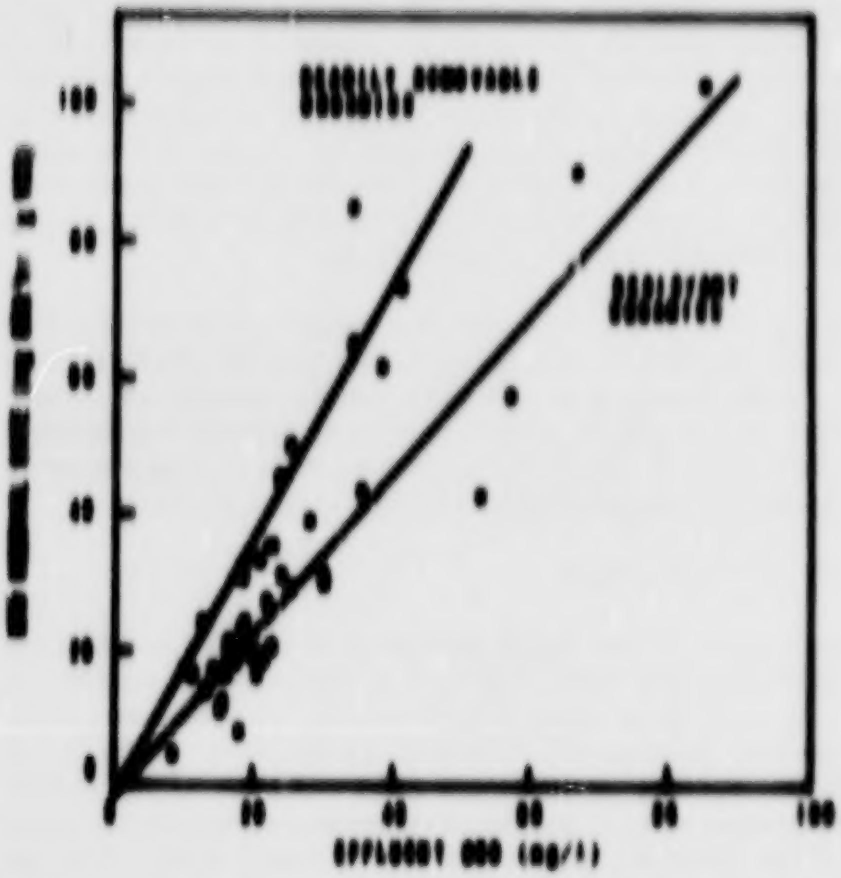
where:

$$L_m = \frac{L_o + R L_e}{1 + R}$$

r = BOD removal rate constant, min^{-1}
L_m = BOD of wastewater after mixture of raw wastewater or primary effluent with sludge recycle, mg/l
L_e = Clarifier effluent BOD, mg/l
t_a = Aeration tank detention time, minutes (including recycle)
L_o = Raw wastewater or primary effluent BOD, mg/l
R = Sludge recycle as percent of influent flow
S_a = MLVSS, mg/l

FIGURE 5-9

BOD REMOVAL CHARACTERISTICS FOR VARIOUS COMPLETE MIX ACTIVATED SLUDGE PLANTS (41)



The aeration tank detention time is related to process efficiency (E, percent) and BOD removal rate (r) by the following equation:

$$t_a = \frac{S_a}{100} E \times \frac{1}{r}$$

where:

$$E = \frac{S_a - S_e}{S_a} \times 100$$

Operating data from over 80 plants with various activated sludge modifications were analyzed using the Weston procedure to determine BOD removal rate constants. Operating data for this analysis were taken from references (3)(4)(8)(12)(13)(15)(16)(21)(22)(23)(28)(29)(37)(38)(40)(44)(45)(46)(47). The results are summarized on Figure 5-10. It should be pointed out that the BOD removal rate curves represent only average kinetics with no temperature correction applied, and the various loading ratios were determined using MLSS, not MLVSS. For these reasons, these curves are not recommended for design purposes, but are included merely to illustrate the relative kinetic rates of the modifications. The presence of significant quantities of industrial wastes, which may have different removal rate characteristics, would modify or displace the curves shown.

The BOD removal rate curve for the second stage of a two-stage process on Figure 5-10 represents data obtained from three two-stage biological treatment plants (28)(29)(45). Two of these plants use activated sludge as the first stage, while the third plant uses trickling filtration. The BOD removal rates for second-stage treatment are markedly lower than those of conventional activated sludge because the organics remaining in the first-stage effluent are more resistant to biological degradation than those entering a conventional plant.

5.5.3 Oxygen and Air Requirements

Table 5-9 contains ranges for oxygen and air required per lb of BOD removed for the activated sludge modifications previously discussed. These values represent overall process requirements. The total amount of oxygen required will vary within the ranges shown depending upon the F/M, increasing as F/M decreases. In diffused air systems, the air requirements will vary depending on the oxygen transfer efficiency of the diffusers employed. The designer must recognize that the various process modifications will require different air and oxygen distribution patterns. Also, the values shown in the table do not include allowances for nitrification or for other plant air requirements.

FIGURE 5-10

RELATIONSHIP BETWEEN B.O.D. REMOVAL RATE CONSTANTS AND LOADING RATIOS FOR THE ACTIVATED SLUDGE MODIFICATIONS

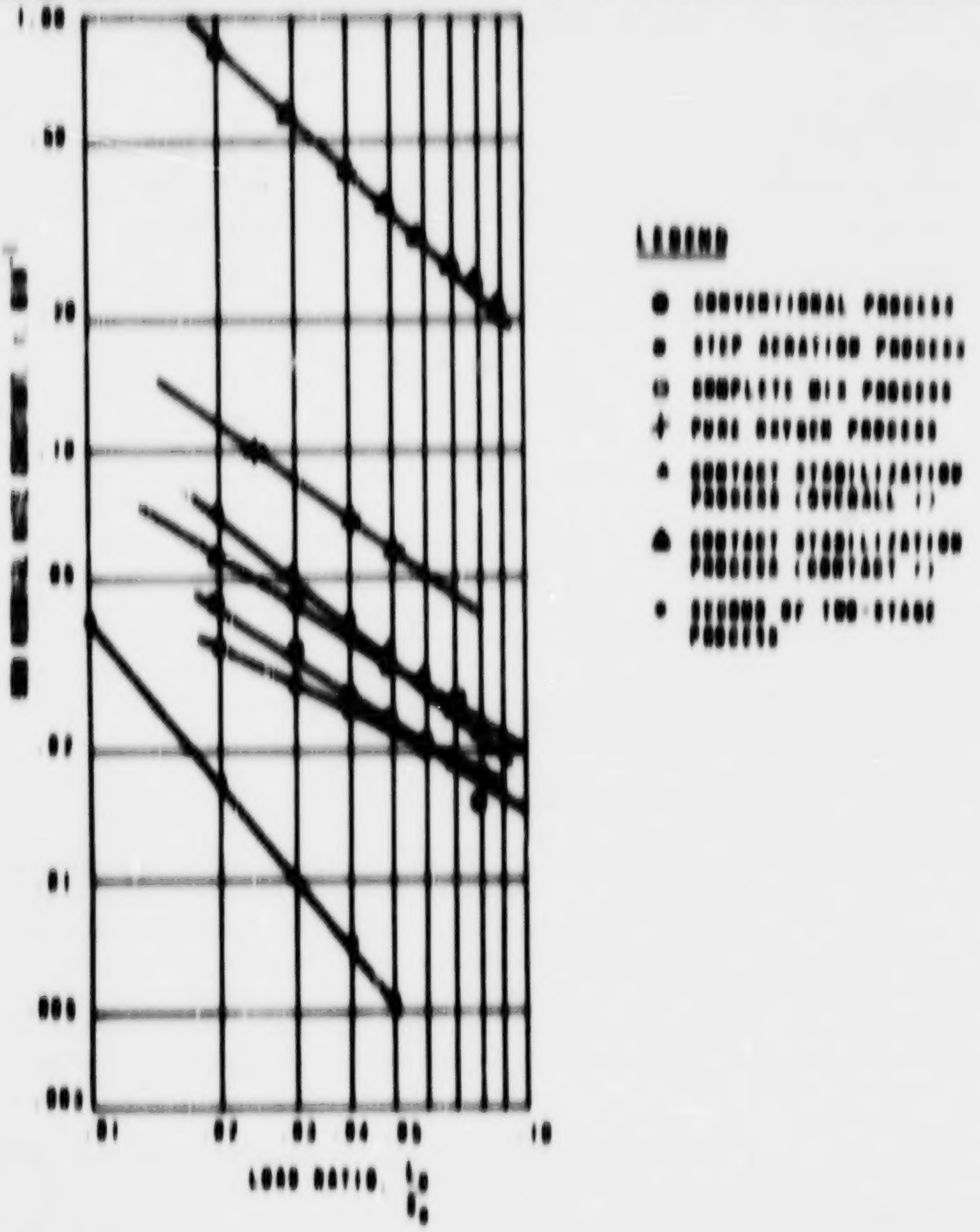


TABLE 5-9
OXYGEN AND AIR REQUIREMENTS FOR ACTIVATED SLUDGE MODIFICATIONS

Process	lb O_2/lb BOD Removed	Standard cu ft Air/lb BOD Removed
Conventional	0.8 – 1.1	800 – 1,500
Step Aeration	0.7 – 1.0	500 – 1,000
Contact Stabilization	0.7 – 1.0	500 – 1,000
Complete Mix	0.7 – 1.0	500 – 1,000
Modified Aeration	0.4 – 0.6	400 – 800
Pure Oxygen	0.8 – 1.4	–

5.5.3 Sludge Production

Normally, the activated sludge processes generate excess sludge in relation to the F/M maintained in the system (48). For normal ranges of F/M (0.2 to 0.5 lb BOD/day/lb MLVSS), the quantity of excess sludge produced in air activated sludge systems varies between 0.5 to 0.7 lb VSS/lb BOD removed (31)(48)(60). The quantity of sludge increases with increasing F/M, since at higher organic loadings less auto-oxidation occurs. The quantity of sludge generated is also temperature dependent, decreasing with increasing wastewater temperature. Data reported by various studies indicate that excess sludge production in pure oxygen activated sludge systems may be slightly less than in air activated sludge systems when operating at similar F/M's (31)(48)(60).

5.5.4 Oxygen Transfer Rates in Wastewater

Oxygen transfer rates in wastewater are affected by various physical and chemical parameters, e.g., temperature, degree of mixing, liquid depth in the aeration tank, oxygen composition of aerating gas, type of aeration device, operating DO, barometric pressure and chemical characteristics of the wastewater. The major area often overlooked in the past by design engineers has been the effect of an industrial waste on the overall oxygen transfer rate of a system. Where the industrial waste makes up a large proportion of the total flow, it is desirable to verify oxygen transfer rates in the laboratory for proper sizing of aeration units.

Oxygen transfer capability for several aeration systems are indicated in Table 5-10. The transfer capability for each system at standard conditions are approximate and may vary somewhat for different makes and models. The effective transfer rates shown are for the specific field conditions listed in the table footnote, and should not be used for design purposes. The relative effective transfer rates will vary considerably for different field conditions, and must be determined for the design conditions at the specific installation.

TABLE 5-10
OXYGEN TRANSFER CAPABILITIES OF VARIOUS AERATION SYSTEMS

Type of Aeration System	Standard Transfer Rate[1] lb O_2/hp-hr	Effective Transfer Rate[2] lb O_2/hp-hr
Diffused Air, Fine Bubble	2.5	1.4
Diffused Air, Coarse Bubble	1.5	0.9
Mechanical Surface Aeration, Vertical Shaft	2.8	1.6
Agitator Sparger System	2.1	1.3
Pure Oxygen (HPO)		
Mechanical Surface Aeration + Cryogenic Generation		3.0
Mechanical Surface Aeration + PSA Generation		1.9
Agitator Sparger System + Cryogenic System		2.8

[1] Transfer Rate at standard conditions, i.e., tap water, 20 deg C, 760 mm barometric pressure and initial DO = 0 mg/l

[2] Transfer rate at following specific field conditions

α = 0.85
β = 0.9
t = 15 deg C
Altitude = 500 ft
Operating DO = 2 mg/l for air aeration, 6 mg/l for oxygenation

5.5.5 Nutrient Requirements

It is necessary that sufficient nitrogen and phosphorus be present in a wastewater so that neither nutrient becomes the limiting factor in microbial growth reactions encountered in the activated sludge process. Normally, supplemental nutrients are not required for municipal wastewater treatment plants because adequate quantities are available in domestic wastewaters to make organic carbon the limiting macronutrient. For optimum operation of the activated sludge process, the minimum ratios of raw wastewater N:BOD and P:BOD are 5:100 and 1:100, respectively (31).

5.5.6 Separation and Return of Activated Sludge

In the past, the importance of the final clarifier as an integral unit of the activated sludge process has not been fully recognized. Improperly designed final clarifiers have resulted in inefficient BOD and SS removals in many activated sludge plants.

Doubt has been raised as to the advisability of designing clarifier overflow rates solely on the basis of average or nominal design flow. It is recommended that final clarifiers be designed on the basis of maximum daily flow rates. This technique provides greater protection against system solids washout, at the expense of a somewhat larger clarifier. Recent work by Dick (32) also indicates that solids loading and thickening aspects must be considered in the design of final clarifiers. Typical design parameters for final clarifiers are presented in Table 5-2.

Control of sludge recycle and wasting rates is the most important operational tool available to the plant operator for intelligently managing the sludge inventory and maintaining optimum loading conditions and sludge settling characteristics. Therefore, it is extremely important to provide sufficient sludge recycle capacity to give the operator the required operating flexibility to handle the highly variable and fluctuating waste loads characteristic of many plants. Excessive detention of sludge within the final clarifier will result in deterioration of the sludge.

Three techniques which have been used to control sludge recycle are:

1. Automatically varying the recycle flow to maintain a set relationship to influent flow
2. Setting the recycle flow at a constant rate based upon the average daily flow
3. Controlling the recycle pumps with a sludge blanket sensor set to maintain a predetermined blanket level in the final clarifier.

A firm sludge recycle capacity of at least 50 percent of average design flow is recommended for the conventional and step aeration processes; at least 100 percent is recommended for the contact stabilization and complete mix modifications. Firm capacity is defined as the available pumping capacity with the largest pump out of service.

5.6 Pilot Studies

The use of pilot facilities for investigating the upgrading of existing activated sludge plants is strongly indicated, in many cases, to ensure that optimum design parameters are selected. There are two general types of piloting facilities available: batch or continuous-flow systems. Batch studies are used primarily to evaluate treatability. Continuous-flow systems generally range in size from bench-scale to 10-gpm units and are used to generate data for design, such as BOD removal rates, oxygen requirements and sludge production.

5.4.1 Batch Studies

Batch laboratory-scale units are subject to all of the inherent difficulties of biological oxidation systems, with the added magnified complexities of large surface-to-volume ratios, small quantities of sludge mass in the reactor and the undesirable factors associated with slug feeding of wastewater. In spite of these inherent difficulties, batch studies have attractive features in that they afford an economic and efficiently controlled method of determining the biological treatability of a wastewater and developing fundamental information concerning the applicability of various activated sludge modifications.

5.4.2 Continuous-Flow Studies

Use of the continuous-flow system is preferable to obtain design parameters since it approximates the operation of an actual plant, permitting evaluation of the effects of variations in wastewater loading or strength. Continuous-flow units must be used on wastewaters which exhibit biostatic or exert toxic effects to permit development of an acclimated biomass. However, most municipal wastewaters do not exhibit these properties unless there is a significant discharge of untreated industrial wastes. A continuous-flow, bench-scale aeration unit is illustrated on Figure 5-11. The initial step in setting up such a unit is to characterize the wastewater and to design the unit and select the proper loadings accordingly. Frequently, the bench-scale unit is designed as an integral aerator-settler system constructed of plexiglass with a volume of 1 to 10 liters. The aeration and settling compartments are separated by a plexiglass baffle wall which may be adjusted to vary the size of the opening between the two zones. Solids separated in the settling compartment may be returned to the aeration compartment by sloping the bottom of the settling zone toward the aeration compartment or preferably by the use of a small pump. Air is supplied through a porous diffuser at the end of a tube. An effluent overflow tube is located at the desired water level. The aeration unit may be fed either by a small metering pump, or by gravity as illustrated on Figure 5-11. Once equilibrium conditions have been established within the unit, the removal of BOD, nitrogen and phosphorus can be evaluated. SS removal is determined from settling tests and the resulting sludge may be evaluated for dewaterability by using a capillary suction device. An approximate determination of the quantity of sludge produced per unit of BOD removed may also be made during a bench-scale study.

A schematic of a medium-sized (0.1 to 1.0 gpm) continuous-flow pilot unit is shown on Figure 5-12. Basically, the system consists of a wastewater feed tank equipped with a mixer to blend the wastewater prior to feeding the aeration unit and to prevent solids deposition in the feed tank. The wastewater from the feed tank and the recycle sludge from the final clarifier are pumped to the aeration tank using peristaltic-type pumps. The difficulty of

FIGURE 5-11

BENCH-SCALE AERATION UNIT

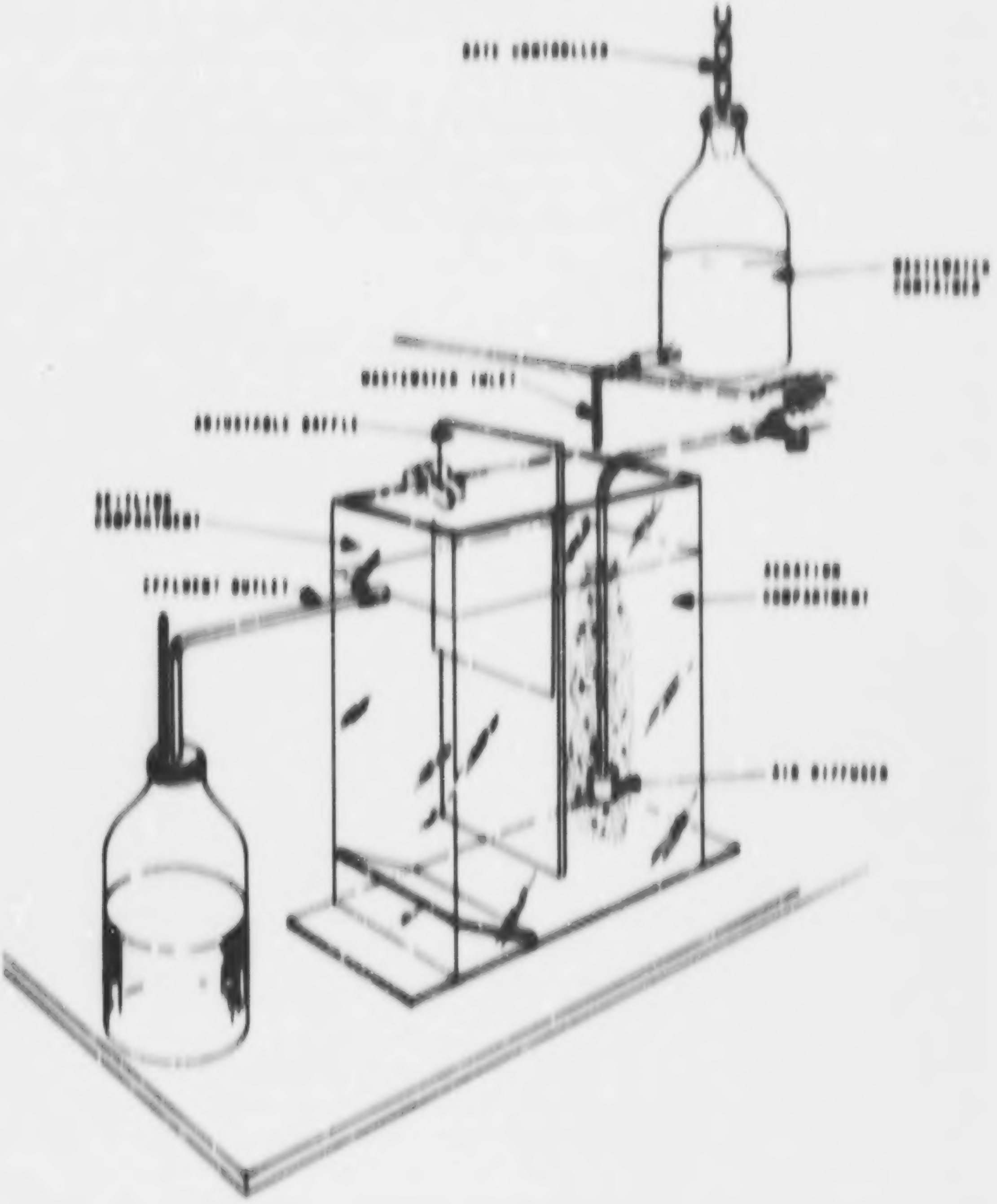
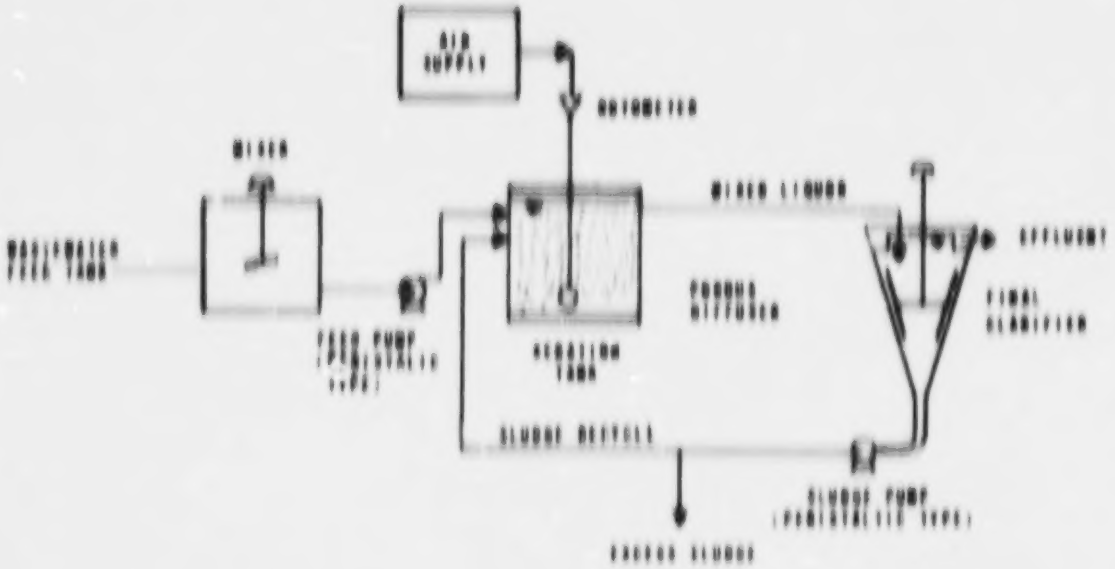

RATE CONTROLLER
WASTEWATER CONTAINER
WASTEWATER INLET
ADJUSTABLE BAFFLE
SETTLING COMPARTMENT
EFFLUENT OUTLET
AERATION COMPARTMENT
AIR DIFFUSER

FIGURE 5.12

SCHEMATIC OF A CONTINUOUS-FLOW PILOT UNIT


AIR SUPPLY
HUMIDIFIER
MIXER
WASTEWATER FEED TANK
FEED PUMP (PERISTALTIC TYPE)
AERATION TANK
MIXED LIQUOR
POROUS DIFFUSER
EFFLUENT
FINAL CLARIFIER
SLUDGE RECYCLE
SLUDGE PUMP (PERISTALTIC TYPE)
EXCESS SLUDGE

pumping extremely low flows makes it essential to provide a reasonably large aerator-clarifier system. Air is normally supplied through porous diffusers controlled by the use of rotameters. The wastewater, after treatment, flows by gravity to a separate clarifier. In the small-scale clarifier, care must be taken to prevent solids deposition on the side walls. The clarifier should have a scraper mechanism to aid in thickening and removal of the mixed liquor solids. If 24-hour composite sampling of feed wastewater and clarifier effluent is required, provision should be made to pump these streams into refrigerated sample bottles.

Two approaches may be applied for the acclimation and growth of a culture of microorganisms for use in a pilot study. An available activated sludge culture may be utilized as the source of microorganisms, with the normal feed to that system being gradually replaced by the wastewater under investigation until satisfactory performance is obtained. Alternatively, culture development can begin with a small quantity of seed organisms and a wastewater feed diluted below the toxicity threshold (if toxicity exists). As the biological mass develops, the toxicity threshold is redetermined and the wastewater concentration is increased accordingly until the culture is capable of handling wastewater at 100 percent concentration. The latter technique is preferred because it provides the best opportunity to observe the growth characteristics of the biological culture as well as potential problems with acute or chronic toxicity.

When the culture is capable of functioning on the undiluted wastewater, data are collected on the performance of the system, beginning with a low feed rate and increasing the feed rate until anticipated design loadings and performance are attained. For various F/M's, the performance and characteristics of the system should be evaluated in terms of:

1. BOD, COD and SS removal
2. Oxygen consumption
3. Waste sludge production
4. Biomass characteristics (microscopic appearance and settling rates)
5. Physical nature of the effluent (turbidity, odor, color, etc.)

5.4.2.1 BOD Removal Rate Determinations

The data collected from continuous flow units can be analyzed using either the Eckenfelder or Weston procedures to define the appropriate BOD removal characteristics for design conditions as previously discussed (41)(42).

8.4.2.2 Oxygen Uptake Requirements

Reliable oxygen consumption data cannot be obtained from pilot plants because of the scale factors involved. However, oxygen uptake studies can and should be conducted on the mixed liquor to obtain energy and endogenous oxygen requirements. A schematic of an oxygen uptake curve for a typical continuous-flow activated sludge pilot unit is shown on Figure 8.13 (81). The slope of the line (m) represents the oxygen required for cell synthesis, while the ordinate intercept (b) represents the oxygen required for endogenous respiration. The net oxygen consumption is expressed by the following equation:

$$O_2 = m\left(\frac{\text{lb BOD removed}}{\text{day}}\right) + b\ (\text{lb VSS under aeration})$$

where:

O_2 = lb oxygen uptake/day
m = energy oxygen, lb oxygen uptake/lb BOD removed
b = endogenous oxygen, lb oxygen uptake/day/lb VSS under aeration

8.4.2.3 Sludge Production

Sludge production in an activated sludge system is expressed as the net effect of two processes as follows:

1. Synthesis of new organisms resulting from the assimilation of the organic material removed
2. Reduction of the weight of organisms under aeration by the process of endogenous respiration

Figure 8.14 is a schematic representation of sludge production from a continuous flow pilot plant (81). The slope of the line (m') represents sludge synthesis, while the ordinate intercept (b') represents the endogenous destruction of solids. The net sludge production is expressed by the following equation:

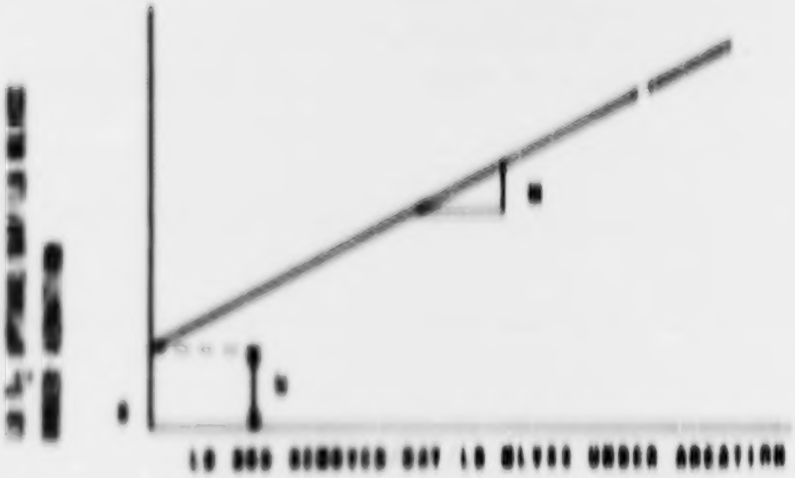
$$\text{VSS produced/day} = m'(\text{lb BOD removed/day}) - b'(\text{lb VSS under aeration})$$

where:

m' = sludge synthesis (lb VSS produced/lb BOD removed)
b' = endogenous destruction of sludge (lb VSS destroyed/day/lb VSS under aeration)

FIGURE 5-18

DETERMINATION OF OXYGEN UPTAKE REQUIREMENTS (51)

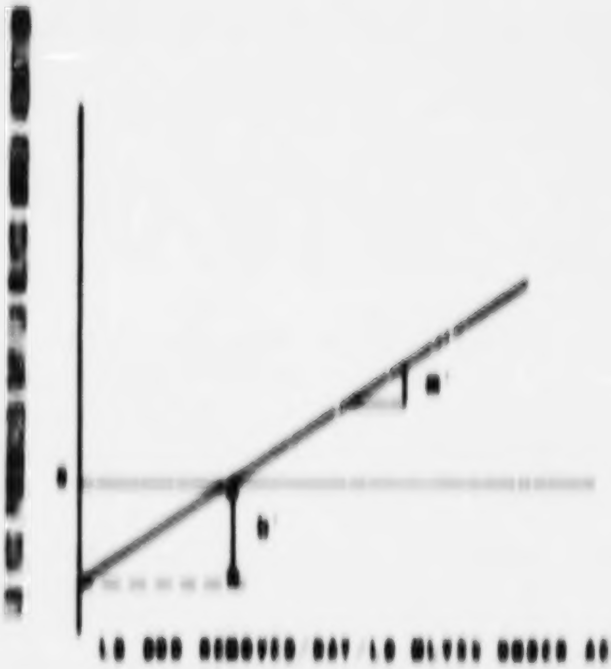


m = ENERGY O_2 (LB O_2/LB BOD REMOVED)

b = ENDOGENOUS O_2 (LB O_2/DAY/LB MLVSS UNDER AERATION)

FIGURE 5.14

DETERMINATION OF SLUDGE PRODUCTION CHARACTERISTICS (51)



m' = SLUDGE SYNTHESIS (LB VSS/LB COD REMOVED)

b' = ENDOGENOUS DESTRUCTION OF SLUDGE (LB VSS/DAY/LB MLVSS UNDER AERATION)

Although the bench-scale and small pilot-scale techniques described above are extremely useful in initial screening and treatability tests, extreme caution must be exercised in extrapolating pilot-generated design data to full-scale plants. The pilot tests should be conducted to approximate the most severe conditions expected for the full-scale plant.

5.8 Activated Sludge Upgrading Techniques and Design Bases

Upgrading activated sludge plants to relieve overloaded conditions, to improve organic removal efficiency, to provide nitrification and to remove nutrients will be covered in the following sections.

5.8.1 Upgrading to Relieve Overloaded Conditions

The following activated sludge modifications are examined as they apply to upgrading existing activated sludge plants:

1. Step aeration
2. Contact stabilization
3. Complete mix
4. Pure oxygen

The step aeration and contact stabilization processes are similar in that both modifications can be incorporated into the upgraded design at a minimum capital investment. The flexibility to operate in both modes is accomplished by sizing the influent step aeration piping so that the entire flow may be introduced in the last bay of the aeration tank, thus permitting operation as a contact stabilization process.

Before examining each individual upgrading procedure, several general statements can be made. Operating data, BOD removal rate constants and volumetric loadings previously discussed indicate that at least one of the activated sludge modifications may be applicable for upgrading an overloaded activated sludge plant. These modifications may require renovation of the air system to supply sufficient air to meet the requirements of the new process because of higher volumetric loadings. The mechanical aerator and agitator-sparger systems are illustrated on Figure 5-2. Even though mechanical aerators afford a high transfer efficiency, their use in an existing basin may pose problems because the geometric configurations required for their most efficient utilization may be quite different than the existing basin configuration. This was found to be true for an upgrading investigation performed for the City of Baltimore, Maryland (55). In an economic comparison it was found that the annual costs for a diffused versus a mechanical aeration system were approximately equal because of the existing configuration of the plug flow basins. However, it was recommended that mechanical aeration be considered for future aeration tank expansion.

Most existing conventional plants use either fine- or coarse-bubble diffused air systems. The fine-bubble system is more efficient, but on the other hand represents a greater capital investment and more costly maintenance problem than a coarse-bubble system. At Milwaukee, Wisconsin, it was found that the type and arrangement of fine-bubble diffusers had a significant effect on oxygen transfer efficiency. The studies showed that ceramic tubes in both spiral and transverse patterns did not perform as efficiently as ridge and furrow pattern ceramic plates located longitudinally. By converting to the ridge and furrow pattern, the overall plant removal efficiency was substantially increased (84).

The agitator-sparger system has an operational advantage over the diffused air unit (coarse or fine-bubble) in that during low flows the air may be reduced but the mixing will be maintained due to the action of the turbine agitator.

5.5.1.1 Upgrading a Conventional Activated Sludge Plant to Step Aeration

Step aeration has been used successfully as an upgrading technique in New York City; Indianapolis, Indiana; and numerous other locations. The case history discussed below illustrates the implementation of this technique at the Wards Island Wastewater Treatment Plant in New York City (85)(86)(87).

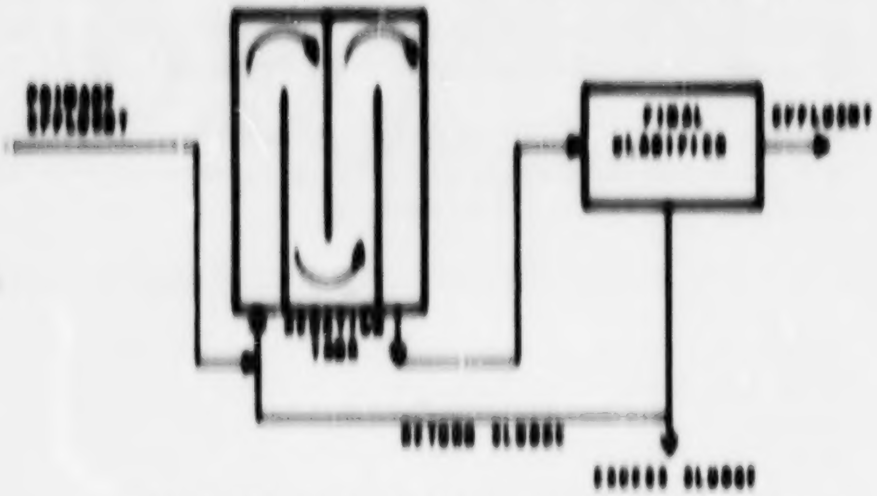
Wards Island was built in 1937 to provide secondary treatment for a design flow of 180 mgd. By 1967, the average flow was in excess of 220 mgd, solids removal performance had become erratic and air consumption was excessive. The problems were remedied and the capacity was increased to 250 mgd by: (1) converting the conventional activated sludge aeration tanks to the step aeration process and (2) modifying the final clarifier inlet-outlet configurations.

The original design provided parallel treatment trains, with an aeration detention time of 7.0 hours and a final clarifier overflow rate of 700 gpd/sq ft at the design flow. The modification of the aeration tank influent systems to provide step aeration, as shown on Figure 5-15, increased the plant's organic loading capacity and greatly enhanced operational flexibility. Using the step aeration process, 12 of the original aerators had sufficient capacity for 250 mgd, providing a nominal aeration detention time of 4.0 hours. A study of plant capacity carried out in 1985 concluded that if the existing preliminary tanks were replaced and the four surplus aerators converted into final clarifiers, the plant could be rated for at least 290 mgd.

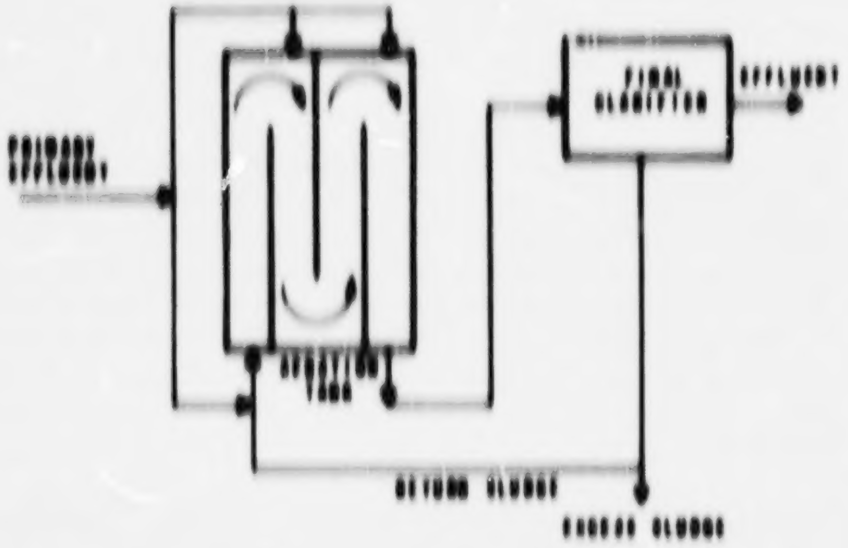
Operating and performance data for the upgraded plant from 1988 are shown in Table 5-11 for one of the four treatment plant batteries (Battery C).

FIGURE 5-15

UPGRADING A CONVENTIONAL ACTIVATED SLUDGE PLANT TO STEP AERATION


PRIMARY EFFLUENT
AERATION TANK
FINAL CLARIFIER
EFFLUENT
RETURN SLUDGE
EXCESS SLUDGE

ACTIVATED SLUDGE SYSTEM BEFORE UPGRADING
(1 OF 10 PARALLEL MODULES)


PRIMARY EFFLUENT
AERATION TANK
FINAL CLARIFIER
EFFLUENT
RETURN SLUDGE
EXCESS SLUDGE

ACTIVATED SLUDGE SYSTEM AFTER UPGRADING
(1 OF 10 PARALLEL MODULES)

TABLE 5-11
OPERATING AND PERFORMANCE DATA FOR THE WARDS ISLAND PLANT, NEW YORK CITY

Description	Operating and Performance After Upgrading (1955) (55)(56)
Flow, mgd	60[1]
Raw Wastewater	
BOD, mg/l	180
SS, mg/l	148
Primary Effluent	
BOD, mg/l	99
SS, mg/l	89
Aeration Tanks	
MLSS, mg/l	1,000
Air Rate, cu ft/gal	0.60
Volumetric Loading, lb BOD/day/1,000 cu ft	46
Organic Loading, lb BOD/day/lb MLSS	0.71
Detention Time, hr	3.4[2]
Final Clarifiers	
Overflow Rate, gpd/sq ft	900
Secondary Effluent	
Overall BOD Removal, percent	95
Overall SS Removal, percent	95
BOD, mg/l	6
SS, mg/l	7

[1] Battery C of 4 batteries.

[2] Excluding sludge recycle.

The costs for the plant modifications projected to an EPA cost index of 170 are as follows:

Final Clarifier Alterations	$ 690,000
Aeration Tank Modifications	310,000
	$1,000,000

5.6.1.3 Upgrading a Conventional Activated Sludge Plant to Contact Stabilization

Contact stabilization has been successfully used as an upgrading technique in Austin, Texas; York, Pennsylvania; and Bergen County, New Jersey. The case history discussed below illustrates the implementation of this technique at Austin, Texas.

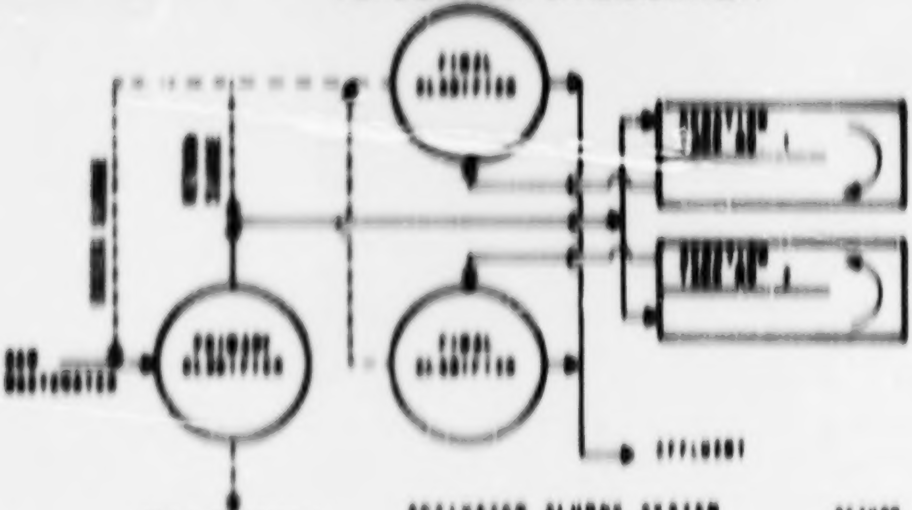
The first full scale application of the contact stabilization concept was developed by Ullrich and Smith (59)(60) at the Austin, Texas Wastewater Treatment Facility. In 1956, the entire plant was converted from the original 6.0 mgd conventional activated sludge operating mode to the "Biosorption" process, later to be renamed contact stabilization. Earlier field studies had indicated that implementation of contact stabilization could effectively control a long-standing sludge bulking problem and significantly increase plant capacity. The cause of the sludge bulking was undetermined, and could be effectively controlled in the conventional operating mode only by bypassing a portion of the primary effluent.

Schematics of the original and upgraded facility are shown on Figure 5-18. Conversion of the plant to contact stabilization was achieved through the splitting of the flow to two separate parallel and independent plants. The two original aeration tanks were split between the two plants, with additional clarification capacity provided. In the upgraded plant, aeration tank A was coupled with the two existing final clarifiers and the original primary clarifier converted to final clarification service. Aeration tank B was matched with two new clarification units. Primary clarification was omitted in the upgraded contact stabilization flow pattern. Table 5-12 compares the operation and performance of the facility before and after upgrading. The data shown for the upgraded plant are for one-half of the treatment facility (Plant A).

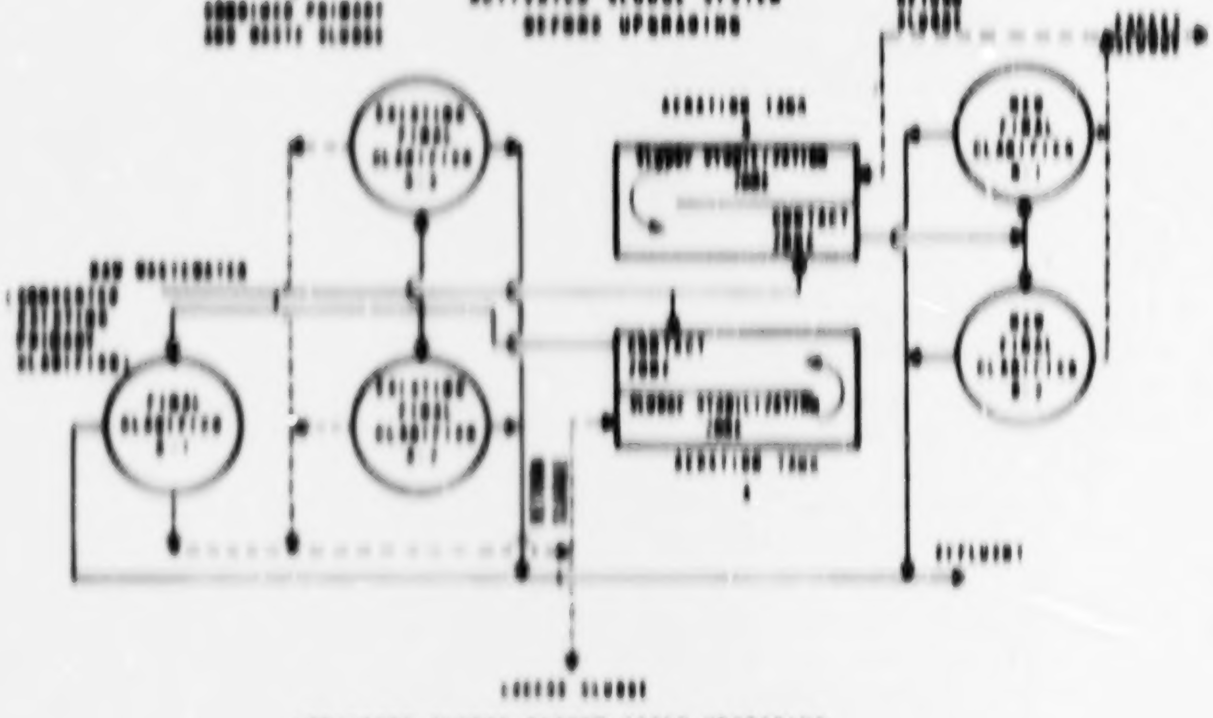
Overall removals averaged 94 percent for BOD and 93 percent for SS in January, 1956. Sludge bulking problems were also eliminated. The data reveal that through conversion to the contact stabilization process and the addition of further final clarification capacity, the upgraded facility could adequately treat over twice the original average daily design flow.

FIGURE 5-16

UPGRADING A CONVENTIONAL ACTIVATED SLUDGE PLANT TO CONTACT STABILIZATION



ACTIVATED SLUDGE SYSTEM BEFORE UPGRADING



ACTIVATED SLUDGE SYSTEM AFTER UPGRADING

TABLE 5-16
OPERATING AND PERFORMANCE DATA FOR AUSTIN, TEXAS

	Overloaded Operating and Performance Data[1]	Upgraded Operating and Performance Data[2] (Plant A)
Flow, mgd	8.2	6.8
Raw Wastewater		
BOD, mg/l	240	250
SS, mg/l	248	247
Primary Treatment		
Primary Effluent BOD, mg/l	151	–
BOD Removal, percent	39	–
Primary Effluent SS, mg/l	98	–
SS Removal, percent	60	–
Aeration Tank		
MLSS, mg/l	1,000	3,333
Sludge Recycle, percent	30.4	56
Air Requirement, cu ft air/lb BOD removed	1,770	776
Volumetric Loading, lb BOD/day/1,000 cu ft	40.9	130[3]
Organic Loading, lb BOD/day/lb MLSS	0.51	0.38[3]
Detention Time in Aerator, minutes[4]	331	59[5]
Detention Time in Stabilization Zone, minutes[6]	–	313
Secondary Clarifier		
Overflow Rate, gpd/sq ft	774	410
Secondary Treatment		
BOD Removal, percent	60	94
SS Removal, percent	70	92
Effluent BOD, mg/l	91	21
Effluent SS, mg/l	73	21

[1] January - September, 1966.
[2] January, 1966.
[3] Including stabilization.
[4] Excluding sludge recycle.
[5] Contact zone.
[6] Based on return sludge flow.

5.5.1.3 Upgrading a Contact Stabilization Plant to Complete Mix Activated Sludge

Past experience with complete mix activated sludge on a large scale has been quite successful, although somewhat limited. Complete mix plants have been installed at Grand Island, Nebraska; Freeport, Illinois; South Tahoe, California; and Albany, Oregon. McKinney (51) and Smith (44) have reported the usefulness of this process for upgrading an overloaded activated sludge plant. When the complete mix process is to be considered as an upgrading technique, the geometric configuration of the existing aeration basin sometimes poses a major problem.

Coralville, Iowa (15) had a contact stabilization package plant which was providing detention times of 2.6 and 6.5 hours, respectively, in the contact and stabilization zones, based on the average influent flow of 667,000 gallons per day. The contact stabilization package plant is shown in plan view on Figure 5-17.

As previously discussed in Section 5.2.5, a contact-zone detention time of this magnitude may result in poor quality effluent because the sludge becomes partially stabilized in this zone and exhibits poorer settling characteristics. After investigation, this was found to be the case in Coralville. Operating and performance data from the plant before upgrading are summarized in Table 5-15. The effluent BOD and SS concentrations were 29 and 34 mg/l, respectively.

To improve the plant's performance, it was decided to modify the flow pattern as indicated on Figure 5-17. The influent piping was modified so that the raw wastewater was evenly distributed into what originally was the stabilization zone. No raw wastewater was introduced into the former contact zone. Mixed liquor in the upgraded system proceeded from the former stabilization zone through the former contact zone to the final clarifier. The return sludge was introduced into the former stabilization zone at one point only. Therefore, the upgrading resulted in a "modified" complete mix flow pattern, with an overall detention time of 9.1 hours at an average influent flow of 667,000 gpd.

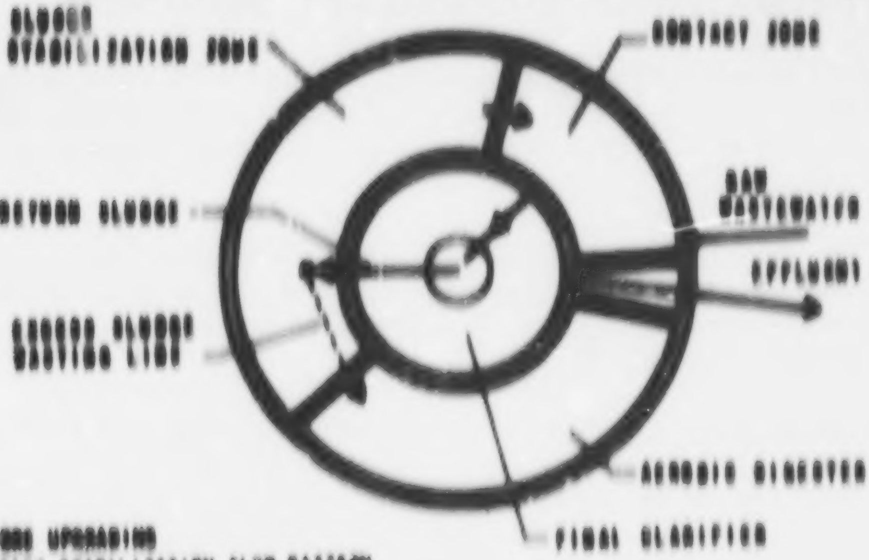
Operating and performance data for the upgraded plant are also included in Table 5-15. The effluent BOD and SS concentrations were lowered to 13 and 6 mg/l, respectively, by the upgrading procedure. The costs associated with this modification are primarily for piping changes. No cost breakdown was available for the modification.

5.5.1.4 Upgrading Existing Treatment Plants to Oxygen-Activated Sludge

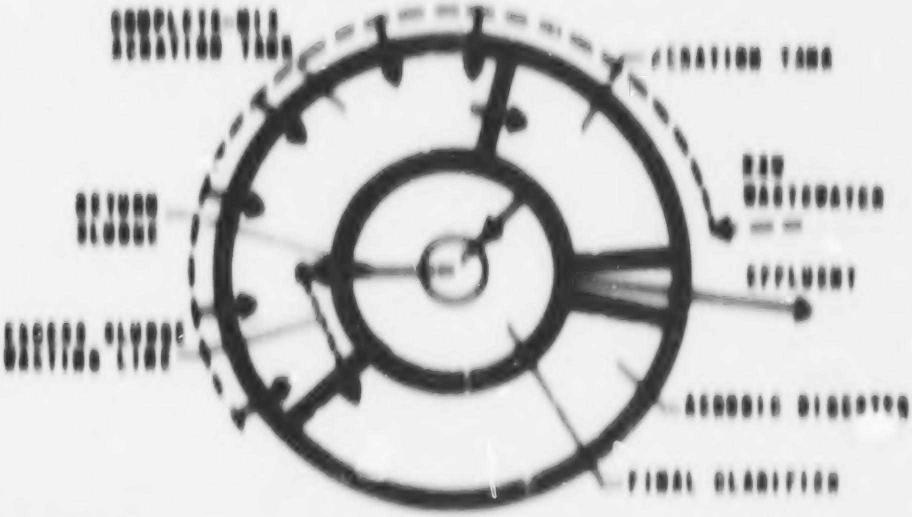
An existing activated sludge system may be upgraded by conversion to a pure oxygen system. However, because of a number of distinct differences between the physical

FIGURE 5-17

UPGRADING A CONTACT STABILIZATION PLANT TO COMPLETE-MIX FLOW PATTERN ACTIVATED SLUDGE (18)


SLUDGE STABILIZATION ZONE
CONTACT ZONE
RAW WASTEWATER
RETURN SLUDGE
EFFLUENT
EXCESS SLUDGE WASTING LINE
AEROBIC DIGESTER
FINAL CLARIFIER

BEFORE UPGRADING
CONTACT STABILIZATION FLOW PATTERN


COMPLETE-MIX AERATION TANK
AERATION TANK
RAW WASTEWATER
RETURN SLUDGE
EFFLUENT
EXCESS SLUDGE WASTING LINE
AEROBIC DIGESTER
FINAL CLARIFIER

AFTER UPGRADING
MODIFIED COMPLETE-MIX FLOW PATTERN

TABLE 5-13
OPERATING AND PERFORMANCE DATA
FOR CORALVILLE, IOWA

Parameter	Contact Stabilization[1]	"Modified" Complete Mix[2]
Average Flow, mgd	0.867	0.867
Raw Wastewater		
BOD, mg/l	185	185
SS, mg/l	150	150
Aeration Tank		
Sludge Recycle, percent	60	60
Contact Zone Volumetric Loading, lb BOD/day/1,000 cu ft	78	–
Contact Zone Organic Loading, lb BOD/day/lb MLVSS	0.4	–
Contact Zone MLSS, mg/l	3,500	–
Complete Mix Volumetric Loading, lb BOD/day/1,000 cu ft	–	22[4]
Final Clarifier		
Overflow Rate[3], gpd/sq ft	750	750
Overall Plant Performance		
BOD Removal, percent	81	90
SS Removal, percent	84	96
Effluent BOD, mg/l	36	18
Effluent SS, mg/l	24	6

[1] Before Upgrading.

[2] After Upgrading.

[3] Based on average flow.

[4] Based on total aeration volume.

components of air and oxygen systems, special consideration must be given to the following:

1. Tank foundations and walls must be checked structurally for increased loadings due to the oxygen dissolution equipment and tank covers.

2. Baffling may be required to sectionalize the existing aeration tank for compatibility with various pure oxygen dissolution configurations.

3. Existing air diffusion piping may have to be removed.

4. Protection must be provided against the potential explosion hazard of pure oxygen or oxygen enriched air.

5. Protection may have to be provided against potential accelerated corrosion due to pure oxygen or oxygen enriched air.

Aspects of using pure oxygen which may make it economically attractive include:

1. Significant increases in volumetric loading may be accommodated in existing aeration tank structures.

2. The oxygen generation equipment may be placed outside and does not require a protective enclosure.

3. Sludge production may be reduced.

4. Expensive renovation of the blower building is eliminated.

5. Liquid oxygen storage may be used to reduce peak energy requirements.

Full scale use of pure oxygen for upgrading municipal treatment plants handling extremely large flows is in the startup stage in Detroit, Michigan, and is under consideration in New York City. Oxygenation at Detroit has been utilized to expand a 800 mgd section of the existing primary treatment plant to secondary treatment. In New York City, a 20 mgd section of the Newtown Creek Wastewater Treatment Plant has been converted to oxygen aeration to upgrade treatment efficiency. This conversion is presented as a case history below (81).

The Newtown Creek Plant has 16 parallel bays, each of which has a design capacity of 20 mgd and consists of an aerated grit chamber, an aeration tank and a final clarifier. The system is designed as a modified aeration process, a process that typically achieves 65 to 70 percent BOD removal when operated at design flow and MLSS concentration. Currently, the Newtown Creek facility is receiving only 55 to 60 percent (11 to 13 mgd/bay) of the design flow. This temporary flow condition has allowed operation of the modified aeration system in a more conventional mode with a nominal aeration detention time of approximately 2.5 hours and an MLSS concentration of 2,000 mg/l, as shown in Table 5.14. When design flow is reached within the next one to two years, operation will have to revert to the modified aeration mode because of air supply limitations. At that time, the City will be confronted with a severe upgrading problem in a land locked neighborhood.

TABLE 5.14
OPERATING AND PERFORMANCE DATA FOR
NEWTOWN CREEK PLANT,
BROOKLYN, NEW YORK

Description	Air[1] System (20)	Oxygen[2] System (31)
Flow, mgd	10.9	20.6
Raw Wastewater		
BOD, mg/l	202	171
SS, mg/l	184	166
Aeration Tank		
MLSS, mg/l	2,000	3,875
Sludge Recycle, percent	20	45
Oxygen Requirements, lb O_2/lb BOD removed	16.1[3]	1.02
Volumetric Loading, lb BOD/day/1,000 cu ft	106	178
Organic Loading, lb BOD/day/lb MLSS	0.92	0.74
Detention Time in Aerator, minutes[4]	150	86
Secondary Clarifier		
Overflow Rate, gpd/sq ft	500	936
Secondary Treatment		
BOD Removal, percent	88	90
SS Removal, percent	80	89
Effluent BOD, mg/l	24	17
Effluent SS, mg/l	37	18

[1] October 1, 1972 through November 30, 1972.
[2] Phase 4 of study, April 8, 1974 through June 2, 1974.
[3] Excludes air diffuser oxygen transfer efficiency.
[4] Excluding sludge recycle.

Pure oxygen is believed to be a good candidate for achieving the required 90 percent BOD removal within the confines of the existing aeration and final clarification tankage at design flow. To evaluate this possibility, one bay of the Newtown Creek Plant was converted to a staged, covered reactor oxygen system, as shown on Figure 5-18. This test system has been operated continuously since May, 1972.

From September 17, 1972 to September 1, 1973, seven phases of a multi-faceted experimental program were conducted. The oxygenation system was evaluated under different loading and diurnal flow patterns in each phase. Operating and performance data for Phase 4 are summarized in Table 5-14. In this phase, the average flow approximated the design flow of the bay, with diurnal fluctuation matching the typical diurnal pattern experienced at the plant. Throughout the seven phases, the pure oxygen system achieved overall BOD removals of at least 87 percent. In Phase 4, the BOD removal equalled the required rate of 90 percent.

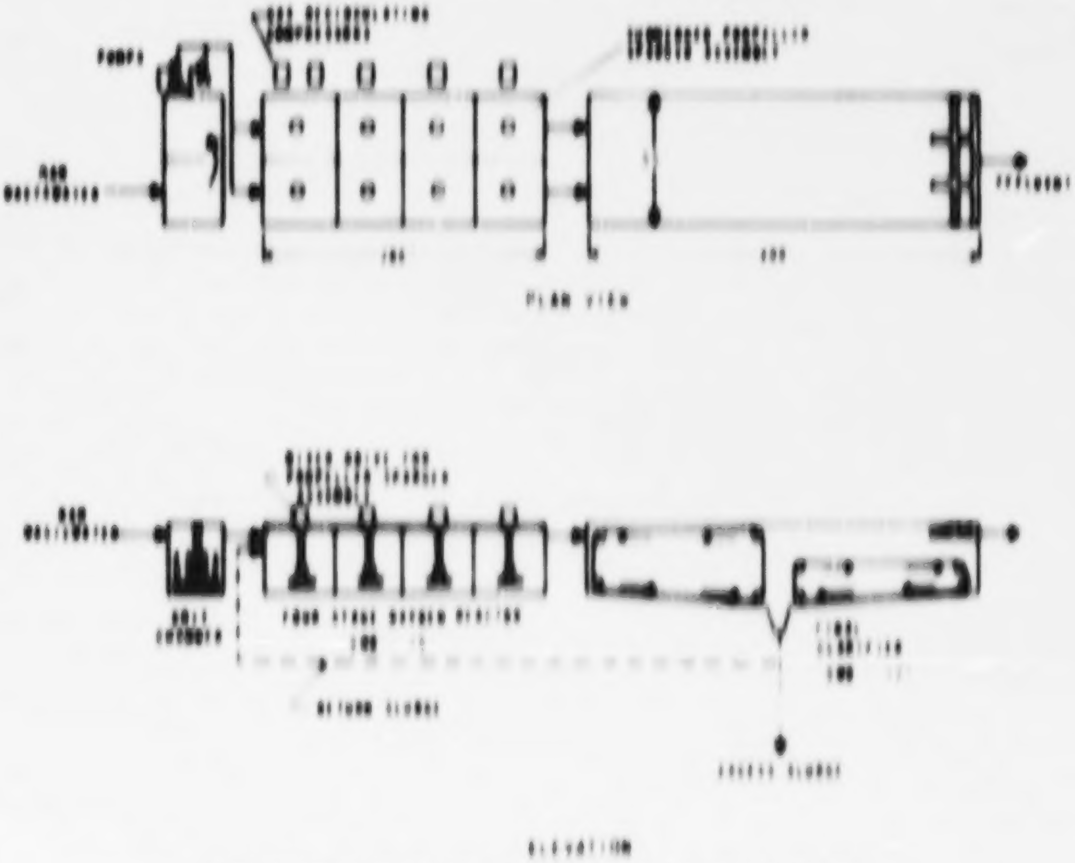
A second case history wherein an existing plant was converted to pure oxygen is described below for Fairfax County, Virginia. In May, 1970, Fairfax County was required to upgrade its Westgate Treatment Plant to prevent the imposition of a building moratorium in the area served by this plant (61). The original plant was designed in 1954 as an 8-mgd capacity facility consisting of two rectangular tanks baffled to form three compartments, as shown on Figure 5-19. The plant was designed to achieve 90 percent BOD removal. By 1970, average flows had increased to 12 mgd and BOD removal was reduced at times to as low as 65 percent.

The Westgate Plant was scheduled to be abandoned by 1975 when its service area is to be incorporated into a regional system tributary to an advanced waste treatment facility. Hence, a minimum cost interim upgrading plan was sought to meet the imposed requirements of 90 percent BOD removal. It was decided to upgrade the facility with pure oxygen activated sludge.

Portions of the existing rectangular tanks were converted to oxygenation reactors. The initial clarification zone was maintained, and the remainder of the tanks was divided longitudinally to form two reactors each, for a total of four. Each reactor was then baffled to form four stages. The first three stages were covered and the fourth left open to the atmosphere. Surface aerators were provided in each stage, but oxygenation occurred in the first three stages only. Two circular final clarifiers were constructed. The upgraded facility is shown on Figure 5-19. Since this was only a short term upgrading, the pure oxygen supply system selected was simply an oxygen storage station in which trucked-in liquid oxygen is stored, vaporized and fed to the reactors. The upgraded facility became operational in October, 1971.

FIGURE 5.18

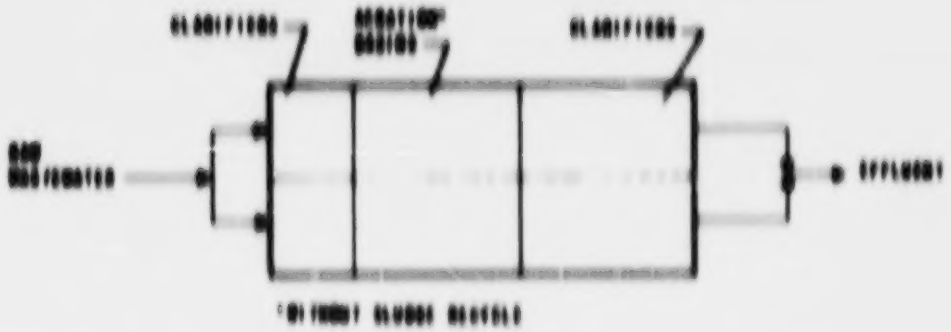
UPGRADING A MODIFIED AERATION ACTIVATED SLUDGE SYSTEM TO OXYGEN ACTIVATED SLUDGE


PUMPS
RAW WASTEWATER
EFFLUENT
PLAN VIEW
RAW WASTEWATER
RETURN SLUDGE
EXCESS SLUDGE
ELEVATION

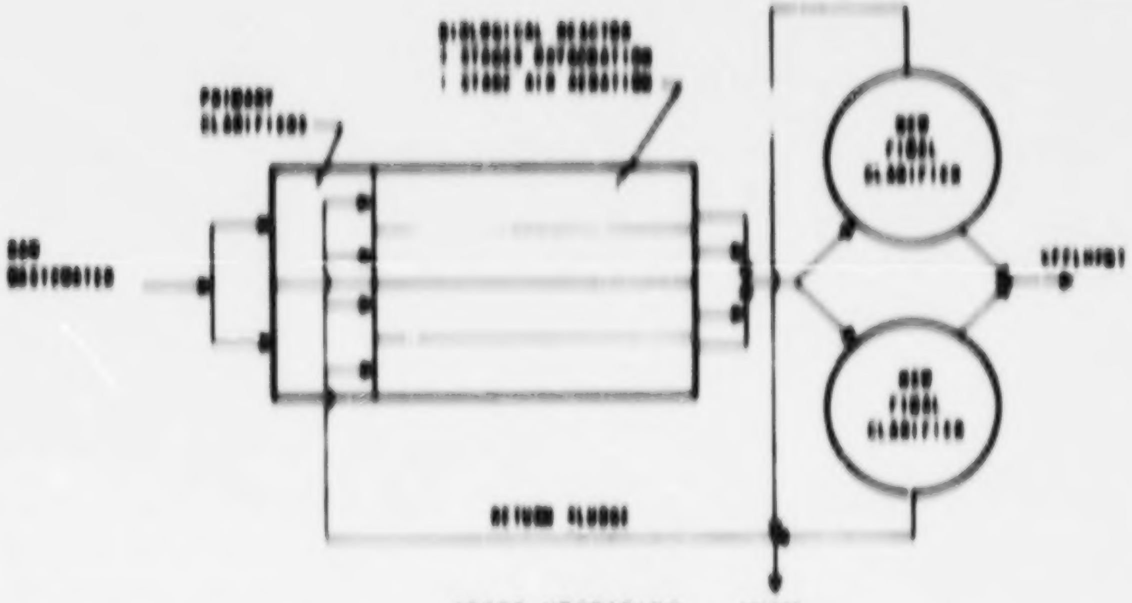
TREATMENT SYSTEM AFTER UPGRADING TO OXYGENATION (31)

FIGURE 5.19

UPGRADING A PRIMARY TREATMENT PLANT TO PURE OXYGEN ACTIVATED SLUDGE TREATMENT


CLARIFIERS
AERATION BASINS
CLARIFIERS
RAW WASTEWATER
EFFLUENT
EFFLUENT SLUDGE RECYCLE

BEFORE UPGRADING


BIOLOGICAL REACTOR
3 STAGES OXYGENATION
1 STAGE AIR AERATION
PRIMARY CLARIFIERS
RAW WASTEWATER
NEW FINAL CLARIFIER
NEW FINAL CLARIFIER
EFFLUENT
RETURN SLUDGE
EXCESS SLUDGE

AFTER UPGRADING

Operating and performance data for the original facility and the upgraded plant are listed in Table 8-15. BOD removals greater than 90 percent are presently being achieved. The cost of upgrading this facility, adjusted to an EPA cost index of 175, was [illegible].

TABLE 8-15
OPERATING AND PERFORMANCE DATA FOR FAIRFAX COUNTY, VIRGINIA

Description	Before Upgrading (1969) (55)	After Upgrading (May, 1972) (55)
Average Flow, mgd	11.5	11.0
Influent BOD, mg/l	240	150
Influent SS, mg/l	–	140
Aeration Tank		
MLVSS, mg/l	–	2,000
Sludge Recycle, percent	–	13.7
Oxygen Requirements, lb O_2/lb BOD Removed	–	0.8-1.00[2]
Volumetric Loading, lb BOD/day/1,000 cu ft	–	115
Organic Loading, lb BOD/day/lb MLVSS	–	0.40
Detention Time in Aerator, minutes[1]	–	180
Secondary Clarifier		
Overflow Rate, gpd/sq ft	–	500
Solids Loading, lb/sq ft/day	–	20.3
Overall Treatment		
BOD Removal, percent	40	91
SS Removal, percent	–	87
Effluent BOD, mg/l	–	12
Effluent SS, mg/l	–	19

[1] Does not include sludge recycle.

[2] The higher uptake rate occurs in warm weather operation due to the higher activity of the biomass.

5.5.3 Upgrading to Improve Organic Removal Efficiency

Upgrading techniques previously discussed relate to the ability of existing activated sludge facilities to handle increased loads and/or overcome operational problems by providing modifications to meet existing effluent standards. However, there may be a need to meet higher effluent standards even though the existing facilities are not overloaded and are performing in accordance with design expectations. Table 5.10 contains suggested alternatives for improving effluent quality under these conditions, along with a range of anticipated improvement in performance for each alternative.

It should be emphasized that, in cases where new unit processes are constructed downstream of existing activated sludge facilities, the improvement in overall organic removal will be a direct function of the BOD removal achieved in the "add-on" process. However, where new unit processes are installed upstream of existing activated sludge units, e.g., the use of a roughing filter, the overall BOD removal may not be increased in direct proportion to the amount achieved by the "add-on" process.

A detailed discussion on polishing lagoons, microscreens, filters, activated carbon and clarifier modifications appears in subsequent chapters. The applicability of these alternatives to individual cases should be evaluated in detail prior to the implementation of a particular upgrading procedure.

5.5.4 Upgrading for Nutrient Control and Removal

With the trend toward more stringent effluent standards, the upgrading of activated sludge plants to achieve phosphorus removal and nitrification and/or nitrogen removal is becoming more frequent. Some of the techniques that may be utilized for this purpose are discussed below. A case history for upgrading an existing plant for combined phosphorus and nitrogen removal is presented in Chapter 10.

5.5.4.1 Phosphorus Removal

One of the most commonly used methods to accomplish phosphorus removal in an activated sludge system is to precipitate the phosphorus by addition of a metallic salt to the secondary system. Metallic salts that have been used include aluminum sulfate, sodium aluminate, ferric chloride, ferric sulfate and ferrous sulfate. The selected salt may be applied to the primary effluent as it enters the aerator, directly to the mixed liquor at a predetermined point in the aerator or to the mixed liquor channel leading to the final clarifier. The optimum addition point should be selected through a combination of laboratory jar testing and comprehensive field testing. A flexible chemical addition design permitting addition at several points in the secondary system flow pattern is highly desirable. When sufficient metallic salt dosages are utilized, phosphorus removals of 80 to

TABLE [illegible]
UPGRADING TECHNIQUES FOR IMPROVEMENT OF ACTIVATED SLUDGE TREATMENT PLANT EFFICIENCY

Addition Preceding Existing Process	Existing Process	Addition Following Existing Process	Incremental BOD Removal Across the Added or Modified Process percent
	Activated Sludge		
Roughing Trickling Filter (Rock or Synthetic Media)			[illegible]
Chemical Addition to Primary Clarifier			[illegible]
		2nd Stage Activated Sludge[1]	[illegible]
		Polishing Lagoon	[illegible]
		Multimedia Filter	[illegible]
		Microscreens	[illegible]
		Activated Carbon	[illegible]

[1] A modification if year-round nitrification is required.

90 percent may be obtained. Case histories illustrating the implementation of phosphorus removal in activated sludge plants may be found in the Process Design Manual for Phosphorus Removal (64).

The addition of metallic salts will increase the solids loading on the final clarifiers, and decrease the percentage of volatile solids in the clarifier underflow. If the benefits of chemical addition are to be fully realized, the existing clarifiers must have adequate capacity to handle the higher solids loadings. Also, the existing sludge recycle and processing systems must be capable, respectively, of returning sufficient sludge to the aeration system to maintain proper biological activity in that unit, and to handle the increase in the total quantity of waste sludge generated by the process. If adequate existing capacity is lacking in any of the above, additional construction may be necessary prior to implementation of phosphorus removal.

5.5.5.2 Nitrification

If year round nitrification is a design criteria, the information presented in Table 5.17 may be useful in the preliminary sizing of second-stage process units (65). It should be stressed that the detention time required to achieve nitrification is strongly dependent upon the wastewater temperature and the MLVSS concentration maintained in the system.

TABLE 5.17
DESIGN GUIDELINES FOR NITRIFICATION

Description	Design Parameter
Nitrification Reactor	
Optimum pH range	7.6 to 7.8
Maximum influent BOD, mg/l	40 to 50
Tank configuration	plug flow
MLVSS, mg/l	1,000 to 2,000
DO at average loading, mg/l	2.0
Minimum DO at peak loads, mg/l	1.0
Sludge recirculation, percent	50 to 100
Detention time based on average flow, hr	3 to 6
Oxygen requirements (stoichiometric), lb O_2/lb NH_3-N	4.6
Clarifier	
Maximum allowable overflow rate, gpd/sq ft	1,000

Nitrification reactors should be designed to ensure that plug flow will exist throughout the reactor. This is because the rate of oxidation of ammonia nitrogen is basically linear, requiring full utilization of the available detention time to complete the oxidation to the higher valence form of nitrate nitrogen. Plug flow is probably best ensured by providing a series of compartments within the reactor. Compartmentation has the further advantage that reactors or portions of reactors may be taken out of service when ambient temperatures are such that the full reactor design capacity is not required to accomplish nitrification.

Air requirements for completing the oxidation process can vary widely depending on pH, wastewater temperature, diurnal flow variations and chemical characteristics of the influent to the reactors (66). These changing oxygen requirements can be met with properly monitored and controlled diffused air or mechanical aeration systems, or a combination of the two.

If nitrification is not required during the colder months, a two-stage system may not be necessary and nitrification may be accomplished within the existing single-stage activated sludge system. The degree of nitrification that can be attained in a single-stage activated sludge system depends on the SRT, mixed liquor DO concentration and wastewater temperature (67). The SRT or rate of wasting sludge is of primary importance. If sludge is wasted at too high a rate, the nitrifying bacteria will be washed from the system. Generally, nitrification begins at an SRT of about five days, but does not become appreciable until the SRT reaches about 15 days, depending upon the temperature. The existing single-stage air supply system must be sufficient to provide the additional oxygen needed to oxidize the ammonia nitrogen.

5.5.4.4 Nitrogen Removal

Denitrification may be added to the nitrification process as the second stage of the nitrogen removal sequence. In this stage, nitrate is reduced to carbon dioxide, water and nitrogen gas following addition of methanol to provide the carbon source. Other methods of removing nitrogen from wastewaters include ammonia stripping, chlorination and ion exchange.

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CHAPTER 6

CLARIFICATION AND CHEMICAL TREATMENT

6.1 Advantages of Upgrading Clarifiers

6.1.1 General

One of the more important considerations in any upgrading situation is the utilization of overloaded clarification units. In many cases, it is possible to achieve substantial improvements in their performance by the use of appropriate upgrading techniques.

By improving the performance of existing clarifiers, the requirement for new facilities can be minimized. In a typical activated sludge plant, it can be expected that with an increased capture of 1.0 mg/l of SS, a concurrent reduction of about 0.6 mg/l of BOD will be attained in the final effluent. Consequently, the required capacity of new units providing either physical or biological treatment may be reduced.

Clarifier performance can be improved in two ways. The first is to improve the physical characteristics of the tanks to achieve more efficient settling and compaction. The second approach is to introduce chemicals into the wastewater to improve the settling characteristics of the SS.

6.1.2 Primary Clarifiers

The upgrading of primary clarifiers has the following advantages:

1. Improved BOD and SS removals at the same flow or similar BOD and SS removals at somewhat higher flows.
2. An increase in quantity of primary sludge produced which can be more readily thickened and dewatered than secondary sludge.
3. A decrease in quantity of secondary sludge produced.
4. A decrease in organic loading to secondary treatment process units.

Primary clarifier performance significantly influences the extent of secondary treatment required and, in most cases, affects the overall effluent quality of existing treatment plants. Also, since clarification is the most economical way to remove suspended and colloidal pollutants, every effort should be made to improve the primary clarification process before additional primary or secondary facilities are considered.

6.1.3 Secondary Clarifiers

The performance of secondary wastewater treatment systems is determined by comparing the quality of the overflow from secondary clarifiers to that of the incoming wastewater. The biological treatment unit converts a portion of the soluble and insoluble organic pollutants to suspended organic solids (biological). Unless these organic solids are effectively removed in the secondary clarifiers, the treatment process cannot be considered a successful operation. In fact, of all the process design variables which can affect overall plant performance, those which are selected for secondary clarification are among the most critical.

Effective secondary clarification also will allow more efficient disinfection, reduce the frequency of cleaning chlorine contact tanks, and provide an aesthetically pleasing effluent.

6.2 Process Design of Clarifiers

6.2.1 Primary Clarifiers

Primary clarifiers are designed mainly on the basis of hydraulic overflow rates. In practice, the designer will select appropriate overflow rates for the average and peak flows being considered. Required surface areas are then calculated for both conditions and the larger area is used. In many cases, especially in combined collection systems, the peak flow conditions will govern the design. It should be recognized, however, that clarifier efficiency at peak flows is a function of both magnitude and duration. Therefore, it is of extreme importance that the designer analyze past flow data and the characteristics of the collection system in order to avoid a design based on a short duration peak which may have little effect on clarifier efficiency.

After selection of an overflow rate, the detention time in the clarifiers will be set by the depth. Generally, SS in raw wastewaters are susceptible to some degree of flocculation. Adequate detention time should be provided to allow this flocculation to occur and in turn increase the clarifier efficiency. Normally, the depth should be set to provide a detention time of 90 to 150 minutes.

For rectangular tank design, the longitudinal velocity of flow must also be considered. Excessive velocities can result in scouring of settleable solids with subsequent losses to the effluent. Peak velocities greater than four to five fpm should be avoided. Tank length to width ratios of less than 3:1 should be avoided to prevent short circuiting.

Typical design parameters for primary clarifiers are presented in Table 6-1. These parameters are based on municipal wastewaters containing pollutants in the concentrations normally encountered. For wastewaters containing a high percentage of industrial wastes, it may be advantageous to initiate a testing program to determine design criteria. Based on the values presented in Table 6-1, it can be anticipated that a well-designed primary clarifier will be capable of removing 30 to 35 percent of the BOD and 50 to 60 percent of the SS from a domestic wastewater.

TABLE 6-1
TYPICAL DESIGN PARAMETERS FOR PRIMARY CLARIFIERS

Type of Treatment	Overflow Rate Average	 Peak	Depth
	gpd/sq ft		ft
Primary Settling Followed by Secondary Treatment	800-1,200	2,000-3,000	10-12
Primary Settling with Waste Activated Sludge Return	600-800	1,200-1,500	12-15

6.2.2 Secondary Clarifiers

The function of secondary clarifiers varies with the method of biological treatment utilized. Clarifiers following trickling filters must effectively separate biological solids sloughed from the filter media. Clarifiers in an activated sludge system serve a dual purpose. In addition to providing a clarified effluent, they must also provide a concentrated source of return sludge for process control. Adequate area and depth must be provided to allow this compaction to occur while avoiding rejection of solids into the tank effluent. Secondary clarifiers in activated sludge systems are also sensitive to sudden changes in flow rates. Therefore, the use of multispeed pumps for in-plant wastewater lift stations is strongly recommended where flow equalization is not provided.

The design of clarifiers following trickling filters is based on hydraulic overflow rates similar to the method described for primary clarifiers. Design overflow rates must include recirculated flow where clarified secondary effluent is used for recirculation. Because the influent SS concentrations are low, tank solids loadings need not be considered. Typical design parameters for clarifiers following trickling filters are presented in Table 6-2.

Clarifiers in activated sludge systems must be designed not only for hydraulic overflow rates but also for solids loading rates. This is due mainly to the need for both clarification and thickening in activated sludge clarifiers to provide both a well clarified effluent and a concentrated return sludge.

When the MLSS concentration is less than about 3,000 mg/l, the clarifier size will normally be governed by hydraulic overflow rates. At higher MLSS values, the ability of the clarifier to thicken solids becomes the governing factor. Therefore, solids loading rates become more critical in determining tank size. Design size should be computed for both average and peak conditions to insure satisfactory effluent quality at all times.

TABLE 6-2
TYPICAL DESIGN PARAMETERS FOR SECONDARY CLARIFIERS

Type of Treatment	Overflow Rate Average	 Peak	Solids Loading[1] Average	 Peak	Depth
	gpd/sq ft		lb solids/day/sq ft		ft
Settling Following Trickling Filtration	400-600	1,000-1,200	–	–	10-12
Settling Following Air Activated Sludge (Excluding Extended Aeration)	400-800	1,000-1,200	20-30	< 50	12-15
Settling Following Extended Aeration	200-400	800	20-30	< 40	12-15
Settling Following Oxygen Activated Sludge with Primary Settling	400-800	1,000-1,200	25-35	< 50	12-15

[1] Allowable solids loadings are generally governed by sludge settling characteristics associated with cold weather operations.

Depth of clarifiers in activated sludge systems is extremely important. The depth must be sufficient to permit the development of a sludge blanket, especially under conditions when the sludge may be bulking. At the same time, the interface of the sludge blanket and the clarified wastewater should be well below the effluent weirs.

For rectangular tanks, criteria similar to those given for primary tanks also apply to secondary tanks. However, in long tanks it is common practice to locate the sludge withdrawal hopper about 1/3 to 1/2 the distance to the end of the tank to reduce the effects of density currents.

Typical design parameters for clarifiers in activated sludge systems treating typical domestic wastewaters are also presented in Table 6-2. The design of these clarifiers should be based upon an evaluation of average and peak overflow rates and solids loadings. That combination of parameters which yields the largest surface area should be used.

6.5 Physical Upgrading of Clarifiers

6.5.1 Optimization of Aeration Tank – Clarifier Relationship

The activated sludge process is flexible in that the operator can vary the rate of sludge wasting, thereby controlling the MLSS concentration to maintain the F/M and SRT at any desired value. However, the MLSS in the aeration tank, in conjunction with the return sludge rate, has a direct effect on the solids loading and capture efficiency of the secondary clarifiers. Therefore when evaluating the performance of an existing treatment plant and/or planned expansions, it is important to analyze aeration tanks, secondary clarifiers, and return and waste sludge pumps as a system.

In many treatment plants, the MLSS concentration will be altered to achieve a desired operating condition in the aeration tanks without considering the possible adverse effects on the secondary clarifiers. It is common practice among many operators to carry high MLSS concentrations to increase SRT, thereby oxidizing more organic matter and reducing waste sludge mass. In many instances, the high solids loading rate associated with high MLSS concentrations will cause the sludge blanket to rise to a level where solids will be swept over the effluent weirs. Thus it is essential that a proper solids balance be maintained between the aeration tank and clarifier.

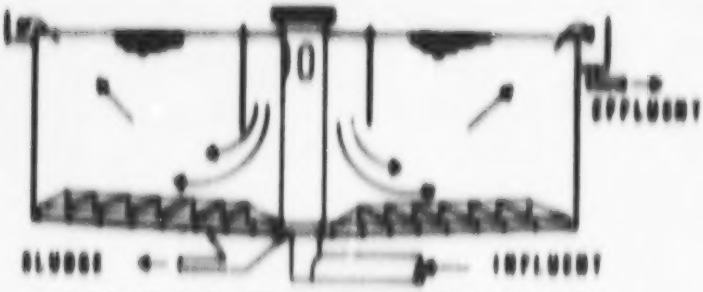
Improved performance through optimization of internal process balances was accomplished at both the Coldwater Creek Wastewater Treatment Plant in St. Louis, Missouri, and the Sioux Falls, South Dakota, Wastewater Treatment Plant. Case histories of the upgrading of these plants presenting the major modifications implemented and resultant removal efficiencies are described in Chapter 13.

6.5.2 Inlet-Outlet Configurations

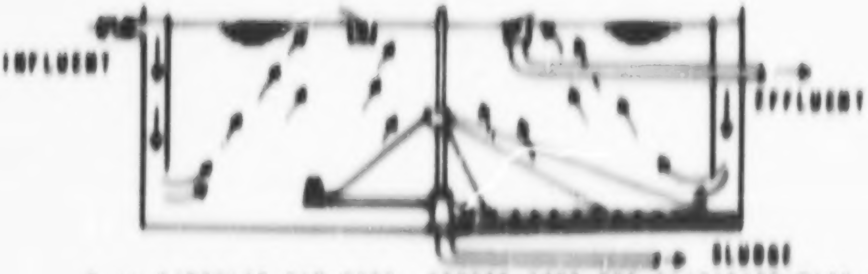
Typical inlet-outlet arrangements for circular clarifiers are shown on Figure 6-1. Figure 6-1a is the most common arrangement currently in use. The influent is carried to the center of the tank by means of a pipe that may be either suspended a few feet below the water surface or constructed beneath the tank floor. A circular baffle is provided at the center of the tank to dissipate the energy generated by the influent discharging into the tank and to distribute flow equally in all directions. The diameter of the influent well should be approximately 15 to 20 percent of the clarifier diameter. The well should extend deep enough into the clarification zone (at least 1/3 the tank depth) to insure good distribution, but should be high enough above the sludge zone to prevent scouring of settled sludge. The clarified effluent leaves the tank over a circular weir located at the perimeter of the tank. For large circular tanks, the clarified effluent may be removed by means of a cantilevered circular effluent trough at a point about 2/3 to 3/4 of the clarifier radius. The purpose of such an arrangement is to reduce weir hydraulic loadings and approach velocities to discourage the creation of currents which may travel along the clarifier floor and up the

FIGURE 6-1

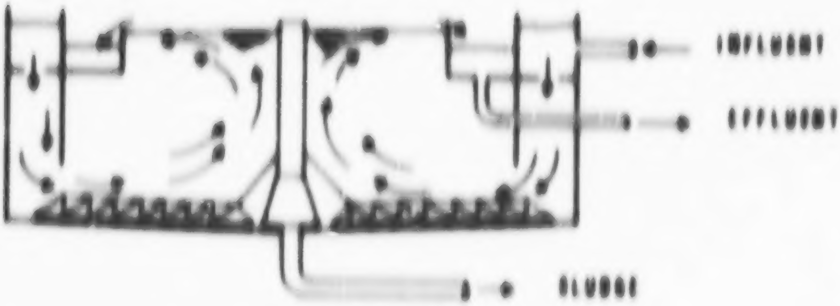
TYPICAL CLARIFIER CONFIGURATIONS


EFFLUENT
SLUDGE
INFLUENT

6-1a CIRCULAR CENTER-FEED CLARIFIER WITH A SCRAPER SLUDGE REMOVAL SYSTEM


INFLUENT
EFFLUENT
SLUDGE

6-1b CIRCULAR RIM-FEED, CENTER TAKE-OFF CLARIFIER WITH A HYDRAULIC SUCTION SLUDGE REMOVAL SYSTEM


INFLUENT
EFFLUENT
SLUDGE

6-1c CIRCULAR RIM-FEED, RIM TAKE-OFF CLARIFIER

clarifier wall. However, many designs of this type have an opposite result when the cantilevered trough is insufficiently far from the tank wall. In such cases, high upward velocities in the annular space between the wall and the trough will actually create a significantly worse condition than that which would have existed with a single peripheral weir.

Rectangular tanks also perform effectively with appropriate inlet and outlet arrangements. Frequently, poor hydraulic distribution in an influent channel common to several rectangular tanks will cause some of the tanks to become overloaded. It is good design practice to provide sufficient head loss in the ports feeding the tanks which will negate the effect of water level variations within the distribution channel.

Inlet baffles for rectangular tanks should extend to at least 1/2 the tank depth to minimize short-circuiting. Effluent weir arrangements should be checked carefully for excessive upward velocities. A common tank deficiency is to have adequate weir length, but with closely spaced troughs creating high rising velocities, enhancing solids losses at higher sludge blanket levels.

Inefficient scum removal can adversely affect plant performance. Scum is usually trapped on the tank surface by means of baffles placed in front of the effluent weirs. Skimmers connected to the arms of the sludge collectors, separate helical devices, or tiltable slotted pipes are used for scum collection and withdrawal. To insure adequate scum capture, baffles should project 9 to 12 inches below the water surface. Since scum removal is ordinarily accompanied by large volumes of water, decanting facilities are desirable to avoid excess water in digestion and dewatering operations.

Two types of rim-feed circular clarifiers are shown on Figures 6-1b and 6-1c. A baffle located a short distance from the tank wall forms an annular space into which the influent wastewater is discharged in a tangential direction. The influent flows spirally around the tank and underneath the baffle. The effluent may be removed by either a centrally located weir trough as shown on Figure 6-1b or by a trough along the tank perimeter as shown on Figure 6-1c. Typically, scum is removed by means of a slotted pipe at the end of the influent annular ring.

It has been demonstrated that the flow pattern generated by the rim-feed clarifier is more stable with varying hydraulic loads than the flow pattern which is developed in circular center-feed clarifiers (1) (2) (3). Bergman (3) noted that rim-feed clarifiers with center takeoff were more stable than center-feed clarifiers when subjected to varying thermal loads, i.e., variation between the temperature of the incoming wastes and the wastes in the clarifier. Studies undertaken by Dague, et al. (2) indicate that center-takeoff rim-feed clarifiers are hydraulically two to four times more efficient than center-feed clarifiers.

Gleasby, et al. (4) indicate that rim-feed and center-feed clarifiers should have the same removal efficiency when handling discrete particles and that rim-feed clarifiers should have higher removal efficiencies when handling flocculent solids.

Although some of the papers cited above suggest that the rim-feed concept is superior to center-feed units, the conversion from center feed to rim feed has not been used extensively as an upgrading technique. Some of the impediments to such a conversion are:

1. Structural modifications to inlet and outlet configurations may be extensive, especially if the inlet pipe passes under the tank floor.

2. Scum removal in rim-feed, center-takeoff clarifiers is somewhat less effective because only the annular influent ring is available for scum separation and removal. With this arrangement, floating sludge is free to escape the tank.

3. The location of the effluent weirs in rim-feed, center-takeoff clarifiers results in relatively short weir length and high approach velocities.

4. The support system of the central weir trough must be designed to avoid interference with existing sludge collector mechanisms.

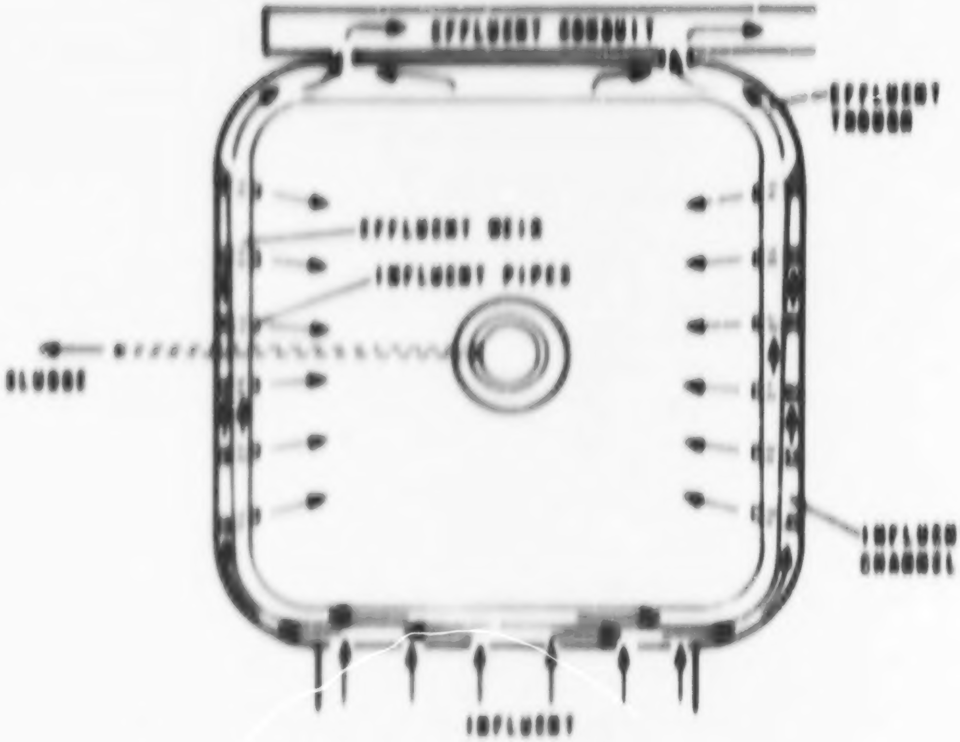
The limitations outlined in the last three comments above are largely eliminated in rim-feed, rim-takeoff designs.

It is now more common to construct center-feed tanks as deep as originally used in rim-feed units (12 to 14 feet). The trend is also towards deeper inlet baffles and to suction-type sludge collectors in center-feed tanks. These features have commonly been installed in rim-feed tanks for many years. Consequently, the advantages of the rim-feed units have been reduced. In most cases, it can be expected that the cost benefit of conversion from center-feed to rim-feed will be lower than the cost benefit of other upgrading techniques.

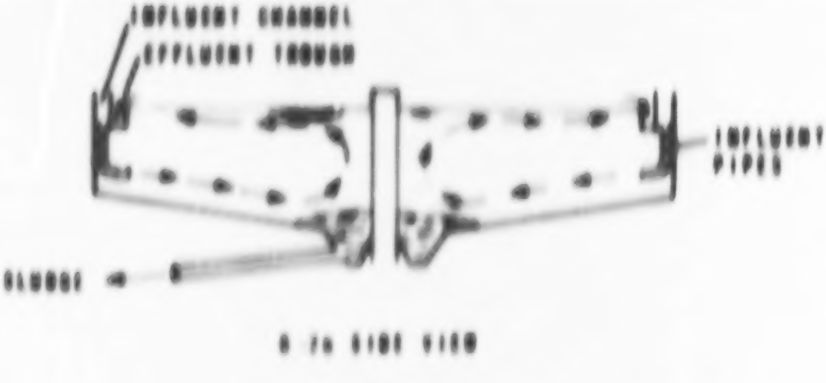
Felt (5) has described a system in the Greater Peoria Sanitary District Sewage Treatment Plant where changes in inlet and outlet designs were implemented to improve clarifier performance. The existing primary clarifiers were square tanks with inlet ports along one side and outlet ports along the opposite side. The secondary clarifiers (following aeration tanks) were similar in design except an effluent weir was provided in lieu of outlet ports. These tanks were experiencing extreme short-circuiting, up to about 1/4 of the theoretical detention time, resulting in poor SS capture. One primary and one final clarifier were modified as indicated on Figure 6-3. After modification, the influent enters the tank from two opposite sides through pipes that discharge near the bottom of the tank. Velocities generated by the influent pipes carried the solids to the center of the tank where the solids are collected. The clarified effluent is removed over a peripheral weir. The modified tanks showed actual and theoretical detention times that were about equal. Peak primary tank

FIGURE 6-2

CLARIFIER MODIFICATIONS AT THE GREATER PEORIA SANITARY DISTRICT SEWAGE PLANT


EFFLUENT CONDUIT
EFFLUENT TROUGH
EFFLUENT WEIR
INFLUENT PIPES
SLUDGE
INFLUENT CHANNEL
INFLUENT

6-2a TOP VIEW


INFLUENT CHANNEL
EFFLUENT TROUGH
INFLUENT PIPES
SLUDGE

6-2b SIDE VIEW

overflow rates as high as 4,100 gpd/sq ft were encountered with very little loss in average SS removal efficiency. Based on these test results, the entire plant was converted to the modified clarifier design. Since this conversion, there have been other process modifications and changes in plant loadings. As a result, it has been difficult to verify the improvement in plant performance due to the clarifier modifications.

6.3.3 Control of Problem Sludges

Two of the major problems encountered in activated sludge systems are rising sludge and bulking sludge. In many instances these problems can be corrected through operational control. Where operational control is limited, plant modifications may be required.

6.3.3.1 Rising Sludges

Sludge accumulation on the bottom of secondary clarifiers for excessively long periods often causes oxygen deficient conditions. If the biological system is nitrifying, the accumulated sludge may undergo denitrification, which occurs when the oxygen contained in nitrites and nitrates is used to satisfy the oxygen demand of the settled sludge. As nitrate-bound oxygen is consumed, nitrogen gas is formed and becomes trapped in the sludge mass. If enough gas is formed, portions of the sludge mass become buoyant and rise to the surface.

The problem can usually be solved by reducing the residence time of the sludge in the clarifier. Corrective actions that may be taken are:

1. Increase the return sludge pumping rate.

2. Increase the speed of the sludge collector (if possible).

3. Increase the sludge wasting rate.

In some instances, floating sludge has been caused by the organism *Nocardia actinomyces*. This condition is difficult to eliminate and is evidenced by a heavy brown scum on aeration tanks and final clarifiers. Some success has been achieved by the following:

1. Chlorination of return sludge or MLSS

2. Lowering DO to 0.5 mg/l or less

3. Reducing MLSS concentration

4. Raising pH to 7.5 to 8.0 by lime addition (6)

Plant modifications that may improve sludge management include increasing the capacity of return and waste sludge pumping, installing sludge blanket finders, and replacing plow-type sludge collectors with suction-type units. The suction type units will remove the sludge directly from the clarifier bottom through suction orifices that sweep the entire bottom of the tank during each revolution. They are recommended for clarifiers with diameters greater than 85 feet.

6.5.5.2 Bulking Sludge

A bulked sludge is one which has poor settling characteristics and poor compactibility. Two principal types of sludge bulking have been identified. One is caused by the growth of filamentous organisms (7) (8) or organisms which can grow in a filamentous form under adverse condition. The other is caused by bound water, in which the bacterial cells composing the floc swell through the addition of water to the point that their density approaches that of water.

The causative factors normally associated with sludge bulking and the recommended preventive actions are as follows:

1. DO content: Limited DO has been noted more than any other cause of bulking. The DO in the aeration tank should be normally maintained at or above 2 mg/l.

2. Process loading: The F/M should be maintained within the recommended range for the operational mode of the type of activated sludge plant being used. Recommended ranges are given in Chapter 5. SRT has also been shown to be an effective tool in preventing the growth of filamentous organisms. Stamberg et al (9) found, for oxygen and conventional activated sludge systems at the EPA-DC Pilot Plant, that for SRT's below five days, both systems exhibited filamentous growth. Such growths did not occur at SRT's greater than five days. When encountering filamentous growth for a few days only, increasing the SRT reestablished a filamentous free sludge in several days.

3. Wastewater characteristics: If it is known that industrial wastewaters are being introduced to the system, the quantity of nitrogen and phosphorus in the wastewater should be checked, since limitations of either are known to favor bulking. Wide fluctuations in BOD and pH should also be checked. The presence of toxic substances has also been implicated as a cause of bulking.

4. Clarifier operation: Evaluation of clarifier performance should be made. This is particularly applicable to rectangular and center-feed circular tanks. Profile sampling of the sludge blanket may show that a large part of the sludge is retained in the tank for many hours rather than the desired 30 minutes. Such a condition may require major modifications to the clarifier's sludge collection system.

When filamentous growths become firmly entrenched in the biomass, it is doubtful any of the above actions will be effective. In these situations, chemical treatment to selectively destroy these growths has been successful as discussed below:

1. Chlorination of wastewater, mixed liquor and return sludge has been found effective in controlling bulking due to filamentous growths. It is ineffective when bulking is due to light, diffuse flocs. Recommended dose rates based on return sludge flow are between 0.2 and 1.0 lb of chlorine per 100 lb of return sludge solids. This method is only considered as a temporary measure and should not exceed 24 hours duration. Selective destruction of filamentous organisms by chlorination generally will cause a turbid effluent until the destroyed organisms are washed out of the system.

2. Application of hydrogen peroxide has also been employed successfully for control of filamentous growths for both air and oxygen activated sludges. At the EPA-DC Pilot Plant, continuous dosing of the return sludge for 24 hours with a hydrogen peroxide dosage of 200 mg/l (based on plant influent) resulted in an immediate improvement in sludge settling and a gradual recovery of system performance (10).

6.3.4 Tube Settlers

The information regarding tube settlers presented in this section is addressed only to the upgrading of existing, conventionally designed clarifiers. A detailed discussion for the design of tube settlers is presented in the Process Design Manual for Suspended Solids Removal (11).

According to the classical theory of discrete particle settling, the efficiency of suspended particle removal in a sedimentation basin is solely a function of overflow rate and is independent of depth and detention time. If the above theory is applicable to raw wastewater or activated sludge floc settling, then the clarifier performance could be improved by introducing a number of trays or tubes in the existing clarifiers. This concept has led to the development of the tube settler. Although tube settlers have been installed in several existing plants, little comparative data have been published to verify their effectiveness as an upgrading technique. However, it can be expected that as additional operational data are developed, the use of tube settlers may warrant further consideration.

Tube settlers have been used in primary and secondary clarifiers to improve performance as well as to increase throughput in existing clarifiers. Conley and Slechta (12) and Hansen, et al. (13) have described the performance of several plant-scale installations of tube settlers in primary and secondary clarifiers. The results of their studies indicate that the overflow rates in primary clarifiers can be substantially increased while producing the same quality effluent as the control unit without the settlers.

One manufacturer of tube settlers has suggested the peak overflow rate for circular secondary clarifiers in activated sludge systems be limited to about 1.0 gpm/sq ft (1,440 gpd/sq ft) for the tube settler area (14). These values apply to 70 deg F wastewaters and should be reduced by a factor of two at 60 deg F. This value is essentially the same as that recommended for secondary clarification without tube settlers. Therefore, tube settlers are not expected to accomplish a saving in tank area. A minimum tank depth of 10 to 12 feet is necessary to accommodate the tube modules and to provide sufficiently low approach velocities.

Tube settlers enhance the ability to capture settleable solids at high overflow rates because the depth of settling has been reduced to a few inches in the tube. It should be realized that tube settlers do not improve the efficiency of primary clarifiers that are already achieving very high (40 to 60 percent) removals of SS. Moreover, tube settlers will neither remove colloidal solids that remain in suspension nor induce additional coagulation to effect added particle removal. A portable tube settler unit has been evaluated following primary clarifiers that were handling paper mill wastewaters which contain a high concentration of solids. With this arrangement, it was possible to appreciably increase the hydraulic capacity of the primary system without reducing the quality of the effluent (15).

The characteristics of the existing tanks will play a large part in determining the feasibility of installing tube settlers. In many upgrading situations, the tube settlers are installed over a portion of the basin near the outlet. For example, in circular basins, the tubes are placed in pie-shaped segments in a ring around the basin outer wall. Major physical factors that must be considered are:

1. Provision of adequate tank depth

2. Provision for adequate sludge management and storage

3. Modifications to existing effluent weir troughs

4. Potential conflicts with sludge and scum collection equipment.

Fouling due to biological growths on the tubes has been cited as an operating problem. This problem has largely been overcome by the installation of permanent air headers beneath the entrance to the tubes. Typical installation details are shown in the Process Design Manual for Solids Removal (11). Algal growths near the water surface can sometimes be controlled by covering the clarifiers to block direct sunlight.

Since the flocculating and settling characteristics of sludge vary from plant to plant, each case should be evaluated separately for suitable design criteria. Small pilot units are available from the manufacturer for this purpose.

Operational data for tube settlers at three locations are presented in Table 6-3. These installations employ tube settlers for clarifying activated sludge mixed liquor. At the two locations, Hopewell Township and Lebanon, where comparative data are available, the tube settlers have markedly outperformed conventional clarification. Case histories for these two installations are presented in the Process Design Manual for Solids Removal (11).

6.4 Chemical Treatment

Chemical addition to primary and secondary clarifiers in this manual is addressed only to increased solids and BOD removal. Chemical treatment for phosphorus removal is covered in detail in the Process Design Manual for Phosphorus Removal (18). A detailed discussion of chemical properties, storage facilities, and feed equipment is presented in the Process Design Manual for Suspended Solids Removal (11).

6.4.1 Chemicals Used

The chemicals commonly used in wastewater treatment are the salts of iron and aluminum, lime, and synthetic organic polyelectrolytes. The iron (ferrous and ferric) and aluminum salts (sodium aluminate or alum) react with the alkalinity and soluble orthophosphate in wastewater to form precipitates of the respective metallic hydroxides or phosphates. In addition, they destabilize the colloidal particles that would otherwise remain in suspension. These precipitates, along with the destabilized colloids, flocculate and settle readily in a clarifier.

Both alum and sodium aluminate exhibit great capability for total phosphorus removal, but the use of alum introduces six times as much dissolved solids to the wastewater as does sodium aluminate (19). Normally, lime is used to precipitate hydrous oxides of iron and aluminum when the alkalinity of wastewaters is low. The reaction of iron and aluminum salts is pH dependent and has to be evaluated for each case to determine the most effective pH range and the optimum chemical dosage.

The addition of lime alone is also effective in coagulating wastewater. The positive calcium ions destabilize colloidal particles while precipitating soluble orthophosphates as hydroxyapatite. Since lime treatment takes place at pH 9.0 to 11.5, pH adjustment may be required before subsequent biological treatment depending on the operating pH and the degree of biologically induced recarbonation that occurs. This requirement may be reduced or eliminated for biological systems that nitrify.

6.4.2 Addition of Chemicals to Primary Clarifiers

In the early stages of wastewater treatment, especially in the late 1930's and early 1940's, chemicals were used to improve the efficiency of primary clarification systems. Later, when these systems were followed by secondary treatment processes, the practice of adding

TABLE 6-3
TUBE SETTLER INSTALLATIONS

Plant Location	Type	Plant Flow Design mgd	Plant Flow Actual mgd	Tube Location	Existing Facility Overflow Rate gpm/sq ft	Existing Facility Effluent SS mg/l	Operational Data Using Tube Settlers: Tube Overflow Rate gpm/sq ft	Operational Data Using Tube Settlers: Tank Overflow Rate gpm/sq ft	Operational Data Using Tube Settlers: Effluent SS mg/l	Reference
Hopewell Township, Pennsylvania	Activated Sludge	0.13	0.13	Secondary Clarifier	0.36	40-70	—	—	—	13
						—	2	0.68	27	
Trenton, Michigan	Activated Sludge	6.5	5.6	Secondary Clarifier	—	—	0.36	0.29	8	16
Lebanon, Ohio	Activated Sludge	0.75	1.25	Secondary Clarifier	0.81	61	0.85	0.62	20	17

chemicals to upgrade primary treatment became unnecessary. However, the technique of adding chemicals to the primary clarifier is still an effective upgrading procedure for a secondary plant when the following conditions exist (30):

1. Wastewater flow is intermittent or varies greatly.

2. Space available for additional clarification facilities is limited.

3. Industrial wastes that would interfere with biological treatment are present.

4. Plant is hydraulically and/or organically overloaded.

5. Improvements in existing treatment performance are required as an interim measure before the addition of new facilities.

When considering the addition of chemicals to primary clarifiers, it is important to examine the effect of increased primary clarifier efficiency on subsequent treatment units. The increased removal of SS and BOD from raw wastewater can affect the downstream biological process in several ways. If the BOD load to the aerator falls below 0.25 to 0.15 lb BOD/lb MLVSS/day for extended periods of time, nitrification conditions can develop in the aerator. This can reduce the total oxygen demand of the effluent, but will impose an added oxygen demand on the aeration facility because the oxidation of one pound of ammonia nitrogen requires about 4.6 pounds of oxygen.

A decrease in loading to the aerator will normally require more careful management of sludge to ensure stable operation of the aeration basin. However, the quantity of excess activated sludge generated under these reduced loading conditions will be substantially less than that generated under normal loading conditions. This may be considered an added advantage of adding chemicals to the primary clarifier, if the additional fines captured chemically do not seriously degrade the quality of the primary sludge.

Little information has been generated regarding the periodic addition of chemicals to the primary clarifier for controlling peak organic or hydraulic loads. This approach, while not always applicable, can frequently be used to maintain system stability during temporary overload.

The effect of polyelectrolyte and coagulant addition on primary clarifier performance is shown in Table 6.4. Where plant information is available, the performance of clarifiers before and after the addition of chemicals is shown. The data show considerable variation in BOD and SS removals. However, the effect of polyelectrolyte addition is pronounced where the existing clarifier performance is poor. The variations shown make it difficult to project the expected improvement due to chemical addition at a specific treatment plant.

TABLE 6-4

EFFECT OF CHEMICAL TREATMENT ON PRIMARY CLARIFIER PERFORMANCE

[illegible]	[illegible]	[illegible]	[illegible]	[illegible]	[illegible]	[illegible]	[illegible]	[illegible]	[illegible]	[illegible]	[illegible]
[illegible]	[illegible]	[illegible]	[illegible]	[illegible]	[illegible]	[illegible]	[illegible]	[illegible]	[illegible]	[illegible]	[illegible]

[illegible]

For proper evaluation, the characteristics of the wastewater and the effects of recycle flows from the sludge processing operations must be known. Such information can be obtained through pilot plant tests or full scale plant trials.

Freese, et al. (37) studied the application of polyelectrolytes for raw wastewater flocculation in the District of Columbia Water Pollution Control Plant. Full scale plant trials were conducted both with and without recycle of elutriate to the primary clarifiers. The overflow from gravity thickening was recycled to the primary tanks for all tests. Without elutriate recycle, primary tank BOD removals were increased from 36 to 45 percent and SS removals from 50 to 61 percent. Overall BOD removal was improved from 74 to 78 percent, but no improvement in overall SS removal was obtained. When the elutriate was added to the primary tank influent without polyelectrolyte addition, poor capture of the elutriate solids occurred. Although capture of elutriate solids improved when polyelectrolyte was added, a major portion of these solids still escaped and were captured in secondary treatment. This led to a gradual accumulation of elutriate solids in the solids processing system, which increased the loadings on thickening, digestion and elutriation tanks. The increased loadings on the elutriation tanks caused by recycled solids caused a gradual decrease in capture efficiency, increasing the solids recycled to primary treatment. It was concluded that polyelectrolyte application to the raw wastewater with elutriate recycle would not prevent the buildup of large quantities of recycled solids through the solids processing components, and that the elutriation tank efficiency would be the focus of further tests.

It has since been found that polyelectrolyte addition to the elutriation tank has increased the capture of solids in the elutriation process from 57 to 92 percent. The resulting improvement in elutriate clarity has enabled the plant to tolerate the recycle without the accumulation of fine solids (38).

Table 6-5 summarizes data of the effect of polyelectrolyte addition in primary clarifiers on the primary and overall plant efficiency. The data reported covers both activated sludge and trickling filter treatment plants.

Schmidt and McKinney (39) studied phosphorus removal by lime addition to the primary clarifier of an activated sludge treatment plant. In this study, the system was operated at a pH value of 9.5 which, during biological treatment, was reduced to a value between 7 and 8. Therefore, no neutralization was required. The lime precipitation step reduced the BOD by 60 percent, SS by 50 percent, and total phosphorus by 80 percent. However, Schmidt and McKinney indicated that the lime-primary sludge was gelatinous in nature and required polyelectrolyte treatment prior to dewatering by vacuum filtration. They further indicated that the mass of primary sludge is about twice that obtained by conventional settling, although the total mass of primary and secondary sludge produced is increased by less than 50 percent. Lime addition to primary clarifiers for phosphorus removal has been used in many locations. In all cases, significant improvements in both SS and BOD removal were noted. Table 6-6 presents the results of some of these studies.

TABLE 6-5
POLYELECTROLYTE ADDITION TO PRIMARY CLARIFIERS

Treatment Process	Coagulant	Dose mg/l	Primary Clarifier: Percent Removal Before Polyelectrolyte Addition BOD	SS	Primary Clarifier: Percent Removal After Polyelectrolyte Addition BOD	SS	Total Plant: Percent Removal Before Polyelectrolyte Addition BOD	SS	Total Plant: Percent Removal After Polyelectrolyte Addition BOD	SS	Reference
Activated Sludge	Purifloc A-21	1	36		60		81		90		37
Trickling Filter	Purifloc A-21	1	23	33	53	76	79	72	85	84	37
Activated Sludge[1]	Purifloc A-23	0.21	31	31	46	51	79	85	81	89	38

[1] Case history presented in Chapter 13

TABLE 6-6
LIME ADDITION TO PRIMARY CLARIFIERS

Location	Lime Added mg/l CaO	Percent Removal in Primary Before Lime Addition BOD	SS	Percent Removal in Primary After Lime Addition BOD	SS	Remarks	Reference
Duluth,	75	50	70	60	75		39
Minnesota	125	55	70	75	90		
Rochester, New York	100	—	—	50	80-90	Jar tests	39
Lebanon, Ohio	145	—	—	66	74	Pilot plant	40
Richmond Hill, Ontario	175	24	47	71	77	Full scale plant	41
Central Contra	370	46	71	74	79	Full scale test	42
Costa, Calif.	403	47	71	69	76	Full scale test	

As mentioned above, the addition of lime to the primary clarifier can be expected to increase the primary sludge mass to about twice that obtained by conventional primary settling, depending on the operating pH and alkalinity of the incoming wastewater. Therefore, a complete evaluation of the sludge handling facilities must be made when considering this technique. For instance, some states have cautioned against this practice when the primary sludge is to be anaerobically digested.

6.4.3 Use of Chemicals in Secondary Processes

Much of the published information available on the addition of chemicals to secondary processes emphasizes their use for phosphorus removal. These studies have shown that in many cases the addition of iron and aluminum salts can significantly improve secondary clarifier performance depending upon the applied dosage, the point of addition, and the flocculent nature of the biomass. The results of several of these studies are presented in Table 6-7.

In certain processes, such as trickling filtration and extended aeration, solids may not flocculate and settle well in the secondary clarifier. In these instances, the addition of iron or aluminum salts will provide a greater benefit in improving plant performance than would a similar chemical addition to an activated sludge that flocculates and settles well.

Polyelectrolytes have also been used to improve the performance of secondary clarifiers. Singer, et al. (48) studied the effect of adding cationic and anionic polyelectrolytes to improve settling characteristics of bulking activated sludge in the laboratory. Their studies indicated that cationic polyelectrolytes at a concentration of 2.0 to 5.0 mg/l were effective in coagulating a bulking activated sludge but that the anionic polyelectrolyte tested had no beneficial effect on improving settling. Goodman and Mikkelson (53) on the basis of full-scale studies, concluded that application of cationic polyelectrolytes to primary clarifier effluent at the rate of 0.1 lb/ton of secondary dry solids increased overall BOD removal efficiency to 95 percent and decreased the loss of solids in the secondary effluent of the activated sludge plant.

Based on studies conducted at the Hanover Treatment Plant by the Metropolitan Sanitary District of Chicago, Zenz and Pivnicka (54) have shown that the addition of alum to aeration tanks (primarily intended for soluble phosphorus removal) improved flocculation of activated sludge. However, their results indicated that increasing amounts of alum floc escaped through the final clarifiers as the dosage of alum increased from an Al:P weight ratio of 1.54 to 1.88. The addition of alum to the aeration tank favored the development of lower organisms, while the higher forms such as protozoa and metazoa were absent. Although higher forms of organisms are adversely affected by the addition of alum, BOD removal is not affected (54) (55). Barth and Ettinger (47) found that dosages of 10 mg/l of alum (as Al^{3+}) in the secondary aerator did not interfere with the nitrification process. Zenz and Pivnicka (54) also indicated that alum sludges can be stabilized anaerobically, and that in this process the precipitated phosphate is not released, and is therefore permanently removed from the system.

TABLE 6-7

EFFECT OF CHEMICAL TREATMENT ON SECONDARY CLARIFIER PERFORMANCE

Derrington, et al. (44) have reported adding 485 gpd of liquid alum (8.3 percent Al_2O_3) to the secondary clarifier of the Richardson, Texas, trickling filter plant treating 1.6 mgd of wastewater. The results indicated a reduction in effluent BOD and SS concentrations from 30 mg/l and 15 mg/l, respectively, to 5 mg/l and 7 mg/l. The effluent phosphorus concentration was reduced from 8 mg/l to 0.5 mg/l. In addition, Derrington, et al. (44) have reported problems of reduced alkalinity in sludge undergoing digestion when alum was used as coagulant in primary treatment at the Richardson, Texas, plant. The addition of alum to raw wastewater was discontinued after nine days total operation to prevent further reduction in alkalinity and pH in the digester. Studies undertaken at Pomona, California (46) indicated that the turbidity of the final effluent may be increased when alum is added to the aeration system. This may occur when the alum dosage is sufficient to lower the pH below the optimum pH for good alum flocculation, resulting in an increase in fines in the final effluent.

Lime addition may not be feasible for upgrading activated sludge secondary clarifiers because of the potential adverse effect of recirculated lime sludge on mixed liquor microbial characteristics. Lime addition to either trickling filter or activated sludge secondary clarifiers will require pH adjustment of the effluent before discharge to the receiving waters. Lime addition to primary clarifiers may be used, if consideration is given to controlling the pH within acceptable limits for the subsequent processes, and to changes in sludge characteristics and handling requirements. Generally, chemical addition will increase the weight of solids and/or the volume of sludge. Accordingly, sludge piping, pumping and process units should be of sufficient size and capacity to handle the increased quantities of sludge.

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CHAPTER 7

EFFLUENT POLISHING TECHNIQUES

7.1 General

The use of effluent polishing techniques is receiving increasing attention as a practical and economical method of upgrading existing secondary treatment facilities to obtain improved organic and SS removal. Effluent polishing is particularly applicable where it is necessary to increase overall efficiency by 10 to 20 percent.

Four effluent polishing processes are considered in this manual: (1) polishing lagoons, (2) microscreening, (3) filtration, and (4) activated carbon adsorption. The reader is referred to the Process Design Manual for Suspended Solids Removal (1) for an in-depth discussion of microscreening and filtration, and to the Process Design Manual for Carbon Adsorption (2) for detailed design information on carbon columns.

7.2 Polishing Lagoons

Polishing lagoons offer an opportunity for increased organic and SS removal at a minimum cost. Both aerobic and facultative lagoons can be used for this purpose.

7.2.1 Aerobic Lagoons

Aerobic lagoons are generally subdivided into two groups:

1. Shallow lagoons with depths ranging from 2.5 to 4.0 feet.

2. Deep lagoons, 7 to 10 feet deep, with aeration devices included to ensure maintenance of aerobic conditions.

The shallow aerobic lagoon is one in which the algae-bacterial symbiotic interrelationship is optimized by providing as much light penetration as possible, and by maximizing photosynthetic efficiency and bacterial oxidation of organic wastes. Operational data from several shallow aerobic lagoons are presented in Table 7-1. The data indicate consistent BOD removals throughout the year, but a marked increase in the concentration of SS in the effluent during the summer periods, when algal activity is at its peak. The seasonal increase in residual SS without a concurrent increase in BOD during the summer months is caused by algae carry-over into the effluent. This indicates that algae present in the effluent do not exert a significant amount of BOD demand during the five-day incubation used in the standard BOD test. The substantial increase in effluent SS during the summer period, however, is a major disadvantage of the shallow lagoon as a dependable year-round polishing technique.

TABLE 7-1

OPERATIONAL DATA FROM SHALLOW AEROBIC POLISHING LAGOONS

Plant Location	Time Period	Detention Time	Flow	BOD In	BOD Out	Lagoon BOD Removal	Surface Organic Loading	SS In	SS Out	Lagoon SS Removal	Remarks	Reference
		days	mgd	mg/l		percent	lb BOD/day/acre	mg/l		percent		
	[illegible]											
[illegible]	Feb. [illegible]	—	[illegible]	[illegible]	[illegible]	[illegible]	[illegible]	[illegible]	[illegible]	[illegible]	Trickling Filter	[illegible]
[illegible]	June [illegible]	—	[illegible]	[illegible]	[illegible]	[illegible]	[illegible]	[illegible]	[illegible]	[illegible]	Effluent	
	Aug. [illegible]	—	[illegible]	[illegible]	[illegible]	[illegible]	[illegible]	[illegible]	[illegible]	[illegible]		
	[illegible]											
[illegible]	June [illegible]	[illegible]	[illegible]	[illegible]	[illegible]	[illegible]	[illegible]	[illegible]	[illegible]	—	Trickling Filter	[illegible]
[illegible]	[illegible]	[illegible]	[illegible]	[illegible]	[illegible]	[illegible]	[illegible]	[illegible]	[illegible]		Effluent	
	Aug. [illegible]	[illegible]	[illegible]	[illegible]	[illegible]	[illegible]	[illegible]	[illegible]	[illegible]	[illegible]	[illegible]	
	Dec. [illegible]	[illegible]	[illegible]	[illegible]	[illegible]	[illegible]	[illegible]	[illegible]	[illegible]	[illegible]	[illegible]	
										[illegible]	[illegible]	
	[illegible]											
	Jan. [illegible]	[illegible]	[illegible]	[illegible]	[illegible]	[illegible]	[illegible]	[illegible]	[illegible]			
	March [illegible]	[illegible]	[illegible]	[illegible]	[illegible]	[illegible]	[illegible]	[illegible]	[illegible]	[illegible]		
	April [illegible]	[illegible]	[illegible]	[illegible]	[illegible]	[illegible]	[illegible]	[illegible]	[illegible]	[illegible]		
[illegible]	Year of [illegible]	—	[illegible]	[illegible]	[illegible]	[illegible]	[illegible]	—	—			
[illegible]										—	Activated Sludge Effluent	[illegible]

[illegible]

An alternative to the shallow lagoon is the deep, aerated, lagoon. These deeper lagoons can operate at greater surface organic loadings than shallow lagoons and yet maintain higher organic removals. Settling of the lagoon effluent is normally necessary to provide efficient SS capture. This can be provided by a quiescent zone at the end of the lagoon, or by a separate clarifier. Since oxygen is supplied to the basin by mechanical devices rather than furnished by the algae-bacterial biosymbiotic relationship, the algae production in the aerated lagoon is minimal compared to the shallow lagoon. Operational data for aerated effluent polishing lagoons (5 to 10 feet deep) are presented in Table 7-2.

TABLE 7-2
REMOVAL EFFICIENCIES FOR DEEP AERATED EFFLUENT POLISHING LAGOONS

Plant Location	Surface Organic Loading lb BOD/day/acre	BOD Removal percent	SS Removal percent	Reference
Washington Borough, N. J. [1]	230	63	79	5
East Windsor Township, N.J. [2]	134	75	75	5
Indian Creek, Kansas [3]	60 to 170	60	50	4

[1] Low Rate Trickling Filter Plant.
[2] Contact Stabilization Plant.
[3] High Rate Trickling Filter Plant.

The lagoon at the Washington Borough Plant has average influent BOD and SS concentrations of 43 mg/l and 70 mg/l, respectively. The average effluent BOD and SS concentrations are 16 mg/l and 15 mg/l, respectively. The East Windsor Plant's polishing lagoon receives BOD and SS concentrations as high as 60 mg/l, while the effluent concentrations are generally about 15 mg/l.

In 1965, the Indian Creek Plant capacity was increased to serve a population of 70,000. Under these conditions it was estimated that the BOD loading to the shallow aerobic lagoon, previously discussed, would increase to 450 lb/day/acre, causing a deterioration of the effluent quality shown in Table 7-1. The shallow lagoon was modified by raising the berm elevation, installing a new effluent structure, deepening the lagoon to 5 feet and providing a perforated hose aeration system. After the above modifications, the lagoon received a flow of 4.0 to 7.0 mgd and provided a detention time of 1.4 to 2.4 days. During 1969, effluent

BOD and SS concentrations averaged 10 mg/l and 9 mg/l, respectively. The highest monthly average effluent BOD and SS concentrations during this time were 17 mg/l and 14 mg/l, respectively.

Deep aerated polishing lagoons may be aerated by either mechanical or diffused air systems. These devices must be designed to provide sufficient oxygen for biological metabolism and adequate mixing. Mechanical surface aerators are commonly used for this purpose in the Southern and temperate climate areas, while diffused air systems have been successful in severe climates where icing may prove troublesome for the mechanical devices.

Eckenfelder (6) has indicated that the power levels required to maintain solids under suspension and to disperse oxygen uniformly throughout the basin are 0.02 to 0.03 hp/1,000 gallons and 0.006 - 0.01 hp/1,000 gallons, respectively.

Eddy (7) studied the degree of mixing provided by mechanical aerators used in treating wastewater from pulp mills. His study indicates that a velocity greater than 0.4 fps should be maintained in the basin to prevent solids deposition, and that mixing energy input varies with the size of the aeration unit. The following values were given as sufficient mixing energy to disperse oxygen uniformly throughout the basin (7):

TABLE 7-1
MECHANICAL MIXING ENERGY REQUIRED FOR OXYGEN DISPERSION

Size of Aerator hp	Mixing Energy hp/1,000 gal
100	0.014
50	0.018
20	0.021

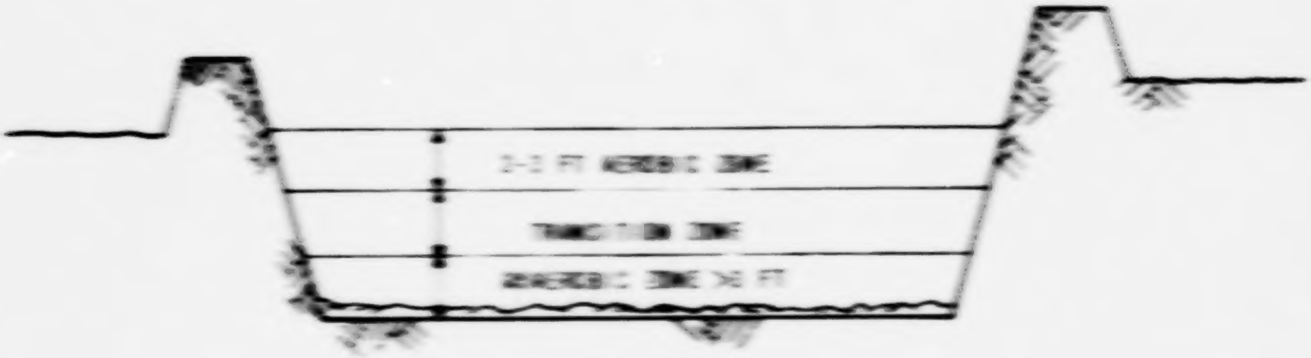
The above discussions indicate that mechanical aerators can be designed to provide either complete mixing of solids including oxygen dispersion, or just to provide uniformly dispersed oxygen. In the latter case, solids deposition will occur in the basin.

7.2.2 Facultative Lagoons

Facultative lagoons are characterized by two distinct zones - aerobic and anaerobic. Hydraulic and organic loadings are such that the DO in the lower section of the lagoon is depleted but an aerobic layer is maintained near the surface. A cross section of a typical facultative lagoon is shown on Figure 7-1.

FIGURE 7-1

TYPICAL CROSS SECTION OF A FACULTATIVE LAGOON


2-3 FT AEROBIC ZONE
TRANSITION ZONE
ANAEROBIC ZONE >3 FT

At Peoria, Illinois, Fall (8) investigated the efficiency of a 10 foot deep polishing lagoon operated for 9 month periods each as a facultative lagoon and as an aerated lagoon. The results of his work are summarized in Table 7.4. It is interesting to note that both the BOD and the SS concentrations in the effluent did not change appreciably during the period when the lagoon was operated aerobically as compared to the facultative operation. Fall also has stated that during the two winters of operation there was no ice on the pond. The lowest temperature of the pond effluent was 40 deg F, and this was recorded after five days during which air temperatures were below zero degrees F. Facultative operation of the lagoon produced small amounts of algae in the pond during the summer period, but no odor problems were noted during the operation of this lagoon.

TABLE 7.4
COMPARISON OF OPERATIONAL DATA FROM FACULTATIVE AND AERATED POLISHING LAGOONS

	Peoria, Illinois		Decatur, Illinois	Springfield, Missouri
Description Type of Secondary Plant	Aerated Activated Sludge	Facultative Activated Sludge	Facultative Trickling Filter	Facultative Activated Sludge
Flow, mgd	0.66	0.65	6.8	18.7
Lagoon Size, acres	0.45	0.45	8.4	10
Average Pond Depth, feet	10	10	5.5	12
Influent BOD, mg/l	58	62	30	83
Effluent BOD, mg/l	34	30	18	30
BOD Removal, percent	41	52	40	64
Influent SS, mg/l	55	55	61	69
Effluent SS, mg/l	17	18	31	26
SS Removal, percent	67	67	49	62
Detention Time, days	1.8	1.82	2.3	1.61
Organic Surface Loading, lb BOD/day/acre	710	747	218	1,292
Air Applied, cu ft air/lb BOD applied	225	0	0	0
Odor	None	None	None	-
Minimum Temperature of Lagoon During Study, deg F	40	40	52	-

Sources: Peoria - Fall (8)
Decatur - Reynolds (9)
Springfield - Hickman (10)

Operational data from the facultative effluent polishing lagoon in Decatur, Illinois, also shown in Table 7-4, indicate that BOD and SS removals averaged 40 percent and 49 percent, respectively, while operating under an organic surface loading of 218 lb BOD/day/acre (9). As seen in Table 7-4, the facultative lagoon at Springfield, Missouri, receives much higher surface BOD loadings (approximately 1,290 lb BOD/day/acre) and still performs creditably, with average BOD and SS removals of 64 and 62 percent, respectively (10).

A major disadvantage of facultative lagoons is the fact that the effluent will have a minimal DO content. Springfield, Missouri, solved this problem by using cascade aeration. The effluent from the polishing lagoon flows over a series of five weirs with a total drop of 75 inches. The average DO in the effluent (September 1970 through March 1971) was 7.0 mg/l with a minimum and maximum, respectively, of 4.0 and 9.9 mg/l (10).

7.3 Microscreening

7.3.1 Principles of Operation

The microscreen is a surface filtration device that has found increasing utility for polishing secondary effluents. It consists of a specially woven polyester or stainless-steel screen mounted on the periphery of a partially submerged, horizontal revolving drum. Influent wastewater enters the drum internally and passes radially outward through the screen, with deposition of solids on the inner surface of the drum screen. A typical microscreen unit is shown on Figure 7-2.

Microscreens are continuously backwashed by water jets located at the top of the drum. These jets normally operate at a pressure of 15 to 50 psig. The backwash water is returned to the head of the plant, and usually totals 4 to 6 percent of the microscreen throughput.

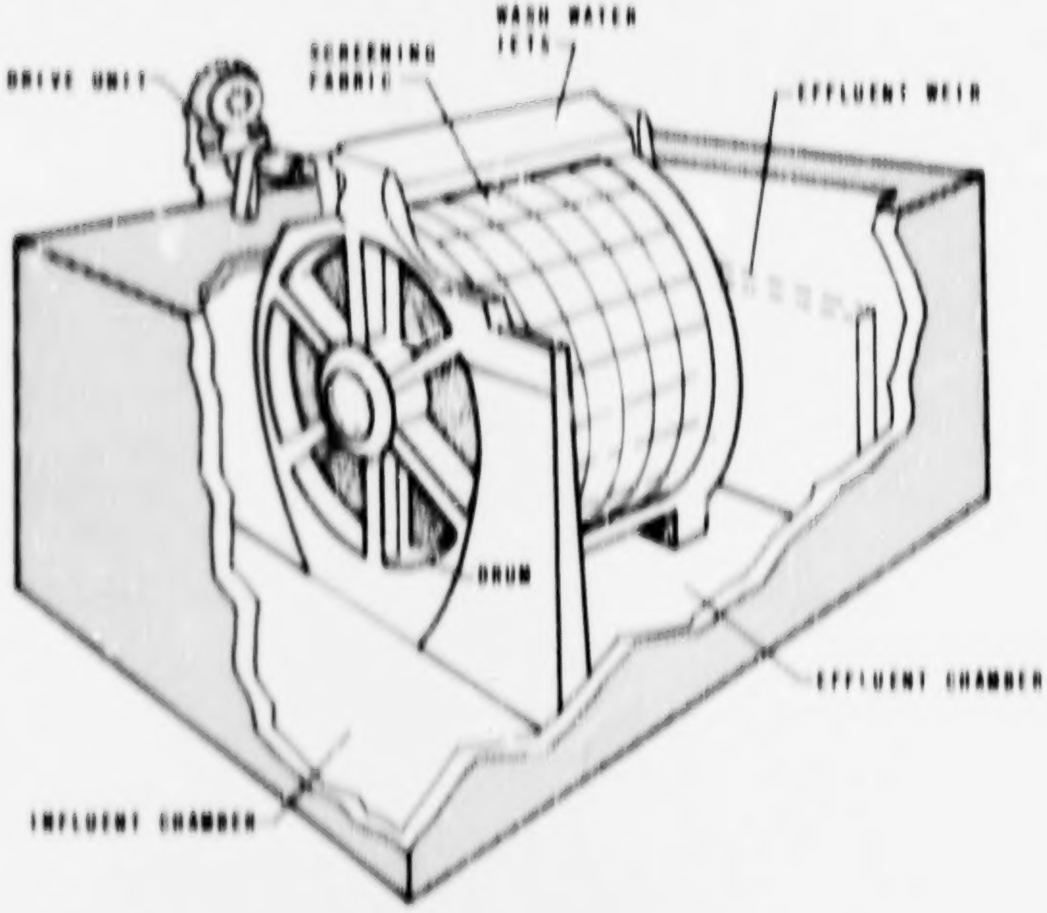
The screens employed have extremely small openings and are available in a variety of sizes, as shown in Table 7-5.

TABLE 7-5
MICROSCREEN FABRIC SIZES

Opening μ	Number of Openings per sq in.
15	[illegible]
23	165,000
35	80,000
60	60,000

FIGURE 7-2

TYPICAL MICROSCREEN UNIT (11)


WASH WATER JETS
SCREENING FABRIC
DRIVE UNIT
EFFLUENT WEIR
DRUM
EFFLUENT CHAMBER
INFLUENT CHAMBER

Typical microscreen drum sizes, capacities, power and space requirements are shown in Table 7-6.

TABLE 7-6
TYPICAL MICROSCREEN POWER AND SPACE REQUIREMENTS

Drum Sizes		Approximate Floor Space		Motors		Approximate Ranges of Capacity
Diameter ft	Length ft	Width ft	Length ft	Drive BHP	Wash Pump BHP	mgd
5.0	1.0	6	6	0.50	1.0	0.07 – 0.15
5.0	3.0	9	11	0.75	3.0	0.2 – 0.4
7.5	5.0	11	16	2.00	5.0	0.5 – 1.0
10.0	10.0	14	22	5.00	7.5	1.5 – 3.0

Courtesy Crane Company

4.0	4.0	7	15	0.75	1.0	0.2 – 0.4
6.0	6.0	10	17	2.00	1.5	0.5 – 1.0
10.0	10.0	14	32	5.00	5.0	1.5 – 3.0

Courtesy Zurn Industries

The weave and shape of individual fabric wires are such that they allow the water from the backwashing jets to penetrate the screen and remove the solids mat which forms on the inside of the screen during its passage through the feed stream. Bodien and Stenburg (11) have noted that only about one half of the applied washwater actually penetrates the screen; the rest flows down the outer perimeter into the effluent chamber. Previously screened effluent can be used as washwater.

As a section of the screen passes through its cycle, the resistance to flow increases as the solid mat forms. Standard design calls for a three-inch head loss at average flow and a six-inch head loss for maximum flows. Newer microscreen models are available with automatic controls to increase drum speed and backwash pressure to accommodate variation in flow and, to a lesser extent, variation in solids loading. Ultraviolet light placed in close proximity to the screen has been somewhat successful in slowing the development of screen clogging slimes. In general, however, the units must be taken out of service on a regular basis (once a month, for example) to have the screens cleaned. In addition, cleaning may be required for removal of iron, manganese, or grease. In cases where oil and grease problems occur, a hot water or steam treatment can be used to remove these materials from the screen.

One of the advantages of using a microscreen for upgrading the performance of existing plants is its low head requirement. It is therefore advantageous to transfer secondary effluent without pumping, to a tertiary microscreening installation to minimize the shear force imparted to the relatively fragile biological floc. Careful design of inlet structures and outlet structures will eliminate turbulent areas and hold the head loss through the entire installation to 12 to 18 inches (12). Typical designs include overflow weirs to bypass part of the flow when the head loss through the screen exceeds a predetermined amount, usually six to eight inches.

7.3.2 Functional Design

Functional design of a microscreen unit involves:

1. Characterization of SS in feed as to concentration and degree of flocculation. The character of the solids affects microscreen capacity, performance and backwashing requirements.

2. Selection of unit design parameter values which will: a) assure capacity to meet maximum hydraulic loadings with critical solids characteristics, b) provide desired performance over the expected range of hydraulic loadings and solids characteristics.

3. Provision of backwash and supplemental cleaning facilities to maintain capacity.

Table 7.7 shows typical microscreen and backwash design parameter values for tertiary solids removal applications.

7.3.3 Performance

Suitable relationships have not been developed for quantitative predictions of microscreen performance from knowledge of influent characteristics and key design parameters.

Where performance must be predicted closely, pilot studies should be made. Where close prediction is less critical, performance data from other locations with generally similar conditions may serve as a guide.

Microscreening has been used for the removal of algae from lagoon effluents. At Bristol, England, algae reductions of 1,565 to 130 algae per ml and 909 to 168 algae per ml were achieved on asterionella, cyclotella and synedra (1).

Many classes of algae (e.g., chlorella) are, however, too small to be removed, even on fine screens (23 μ) and excessive loadings (up to 2×10^6 algae per ml) make this application a limited one.

TABLE 7-7
TYPICAL MICROSCREEN DESIGN PARAMETERS

Item	Typical Value	Remarks
Screen Mesh	20-25 μ	Range 15-60 μ
Submergence	75% of height 66% of area	
Hydraulic Loading	5-10 gpm/sq ft of submerged drum surface area	
Head Loss (H.L.) through Screen	3-6 in.	Maximum under extreme condition: 12-18 in. Typical designs provide for overflow weirs to bypass part of flow when head exceeds 6-8 in.
Peripheral Drum Speed	15 fpm at 3 in. H.L. 125-150 fpm at 6 in. H.L.	Speed varied to control H.L. Extreme maximum speed 150 fpm
Diameter of Drum	10 ft	Use of smaller diameters increases backwash requirements
Backwash Flow and Pressure	2% of throughput at 50 psi 5% of throughput at 15 psi	

Table 7-8 provides performance data for a number of microscreen applications for tertiary solids removal. Reported data indicate the following:

1. Even under best operation, 5 to 10 mg/l residual SS will pass through the microscreen unit.

2. Although the pattern is irregular, performance tends to be better at lower hydraulic loadings.

3. Increases in influent SS are reflected in the effluent but with noticeable damping of peaks.

4. Better removals are obtained with smaller mesh size.

7.4 Filtration

7.4.1 General

Filtration of secondary effluent provides a positive method of SS control, and as such, is one of the more widely used and the most efficient single unit process for upgrading treatment plant performance today. The primary applications of filtration are: (1) direct filtration of secondary effluent, (2) filtration of chemically clarified secondary effluent, and (3) filtration of secondary effluent after in-line chemical injection. Direct filtration has the highest probability of use where existing plants will be required to consistently meet better effluent quality standards than they are presently able to attain.

The major goals of filtration design are:

1. Attainment of required effluent quality
2. Low capital cost
3. Low operating cost.

Present filtration technology provides the designer with a broad selection of filter types and sizes, all of which have been shown capable of producing high quality effluents. For any particular application, adequate consideration must be given to matching filter design and capability to the existing site conditions and wastewater characteristics. Such factors as present and projected plant flows, variability and concentration of filter influent SS, plant hydraulics, and present and projected operation schedules all affect filter selection (11).

TABLE 7.6
MICROSCREEN PERFORMANCE DATA (1)

[illegible]

Low capital costs will generally result from low filter surface areas, and high filtration rates. Operating costs are reduced by designs that minimize backwashing requirements and maximize filter run lengths. In wastewater filtration, run lengths are terminated by either excessive head loss or by deterioration of effluent quality beyond allowable limits. The optimum bed design is one that reaches these limiting conditions at the same time. In some cases, especially in small plants, it may be desirable to design the filter beds so that routine backwashing can be performed once a day.

7.4.2 Filter Types

Filtration systems can be broadly classified as either in-depth or surface. In-depth systems include deep-bed single coarse medium filters as well as dual or tri-media filters. The use of two or more layers of different media having increasing specific gravity in the direction of flow allows gradation of the filter bed from coarse to fine. This allows more efficient utilization of the total bed depth for solids storage than conventionally graded, single medium filters. Multimedia beds normally require more backwash water and higher backwash rates (15-20 gpm/sq ft) than single medium beds (13). Typical multimedia gradations are shown in Table 7-9. The use of several media such as garnet, sand and coal will allow some flexibility in bed design to meet specific effluent quality or desired run lengths.

TABLE 7-9
TYPICAL MULTIMEDIA GRADATIONS (1)

Gradation No.	Garnet Size (Mesh)	Garnet Depth (Inches)	Sand Size (Mesh)	Sand Depth (Inches)	Coal Size (Mesh)	Coal Depth (Inches)
1	40 x 80	3	20 x 40	12	10 x 20	22
2	20 x 40	4	10 x 20	12	10 x 16	15
3	30 x 80	4	20 x 40	9	10 x 20	8

The approach taken in the development of surface filters is to allow filtration to take place on or near the top of relatively shallow (12 in.) single medium filters, and to optimize removal of the accumulated solids. In addition to the standard pressure and gravity surface filters, several innovative techniques for maintaining a continuous clean filter surface have been developed. These include moving bed filters, radial flow filters, radial flow external wash filters, and traveling bed filters.

Although both upflow and downflow modes of filter operation are in use, the downflow filter is far more common for wastewater applications.

As a rule, in-depth filters are better suited to treating strong biological floc, will yield longer run lengths, and are less sensitive to solids loading than are surface filters. Surface filters are better adapted to removing more fragile floc, yield shorter run lengths, and require less backwash water per cycle than in-depth filters.

7.4.3 General Design Considerations

7.4.3.1 Filter Selection and Governing Factors

Performance of different filter systems on a given wastewater may be termed "equivalent" if they produce the same output quality and quantity (1). The major factors that determine overall filter performance are summarized in Table 7-10.

TABLE 7-10
FACTORS GOVERNING FILTER PERFORMANCE

1. **Maximum Available Head Loss**

2. **Filtration Rate**

3. **Influent Characteristics**

   a. SS concentration
   b. Particle size distribution
   c. Floc strength
   d. Temperature (viscosity)
   e. Properties governing adhesion of solids to each other or to media

4. **Media Characteristics**

   a. Grain size
   b. Porosity
   c. Depth
   d. Specific gravity
   e. Configuration

5. **Design of Cleaning System**

   a. Adequacy of cleaning
   b. Washwater volume per cleaning cycle

In general, variations in these factors which increase quality, such as use of finer media, tend to increase head loss and hence reduce output whereas those which increase output tend to reduce quality. Run length may be limited by available head or effluent quality (breakthrough). Washwater requirements depend chiefly on the size gradation, specific gravity of the media, and on the type of supplementary cleaning available.

Another factor that must be examined when selecting a filter for a given wastewater is the effect of chemical pretreatment. Typical situations include activated sludge and trickling filter plants that incorporate alum or iron addition followed by filtration for high level phosphorus removal. In these cases the filter performance may be quite different than for untreated effluents due to the weaker chemical floc normally produced. Generally it is desirable to provide for the addition of polyelectrolytes in installations where aluminum or iron salts are employed for phosphorus removal.

Since influent solids characteristics are variable, the operating head provided in design must cover a range of conditions. If breakthrough consistently requires termination of filter runs at a low head loss, much of the available head will be wasted. On the other hand, if available head is so low that head loss instead of quality limits the run length, filter production capacity will be impaired unnecessarily.

Based on long term practice in water filtration, maximum terminal head loss of 10 to 15 feet has commonly been allowed. In water filtration, this head loss range generally permits runs of 24 or even 48 hours. In wastewater applications, however, heavy solids loadings and high biological floc strength can cause rapid head loss buildup, frequently reaching limiting values well below 24 hours. Filter design should allow sufficient head so that, with maximum influent solids concentrations and floc strengths, minimum run lengths of six hours can be achieved. Also the effect of backwash recycle flows and filter downtime must be considered when sizing units.

Proper cleaning is vital to filter performance. Ineffective cleaning results directly in short filter runs and poor effluent quality and if continued, sets up self perpetuating operational difficulties such as mud balls, slime coating of the media and cracking of the filter bed (14). In wastewater filtration, normal upflow washing plus auxiliary cleaning (air wash or surface wash by water jets) and periodic shock chlorination have proven necessary to maintain filters in proper condition. During a normal backwashing operation, accumulated solids are removed from filters by a rapid upflow of washwater which is returned to the previous treatment units. In small plants, equalization of this flow is necessary to prevent surcharge of upstream treatment units. Washwater sources may include feedwater, filter effluent, or effluent from subsequent treatment units. Washwater storage may be needed if rates required exceed effluent flow available. Backwash rates for most effective cleaning vary with media size, particle density, floc penetration, and strength of floc adherence to the media.

In most applications a number of specific filter designs can provide the required performance. For a given treatment application, if costs and performance variations of alternatives can be defined quantitatively, it is possible to base selection on "least cost." Cost comparisons should include both installed cost and operating cost items such as control required, operating head loss, and backwash water consumption.

7.4.3.2 Pilot Plant Investigations

The extent of pilot investigations which should be conducted as a basis for a given design depends on the size of the proposed installation and on the performance requirements which must be met. For large installations the cost of the necessary pilot plant investigations for optimizing filter design is certainly justified in view of the potential cost savings. When undertaken, these studies should be sufficiently well planned to relate filter performance to the full range of variables expected under actual operating conditions. Figure 7-8 illustrates the type of information that should be developed as a part of an overall pilot investigation to optimize filter design. This figure represents the results of pilot plant studies including only those runs that produced effluents of satisfactory quality. From such data a rational design can be formulated.

Where quality requirements are less stringent, or for very small plants where extensive efforts at cost optimization are not justified, filter design parameters may be selected based upon evaluations of similar installations.

7.4.4 Performance

The single most important factor affecting filter performance is the quality of the secondary effluent produced by the biological treatment. If consistently good performance is exhibited by the biological treatment system, good filter performance can be expected. Conversely, if the biological facility is subject to frequent upsets, filtration will be much more difficult.

Data are presented in Tables 7-11 and 7-12 relating expected filter performance with degree of treatment in activated sludge and trickling filter plants. The data in the tables are based on a filter rate of 5 gpm/sq ft, using a 30-inch deep multimedia filter (55 percent coal, 30 percent sand, and 15 percent garnet) operated to a 10-foot head differential.

Pilot and full-scale performance data for in-depth filters operating on activated sludge and trickling filter plant effluents are shown in Table 7-13. Performance data for surface filters are shown in Table 7-14. As indicated in these tables, both in-depth and surface filters produce an effluent with BOD and SS concentrations generally less than 7 mg/l.

Data describing filter performance at various plants treating chemically treated secondary effluent are shown in Table 7-15.

FIGURE 7-3

TYPICAL PILOT PLANT DATA FOR FILTER DESIGN (13)

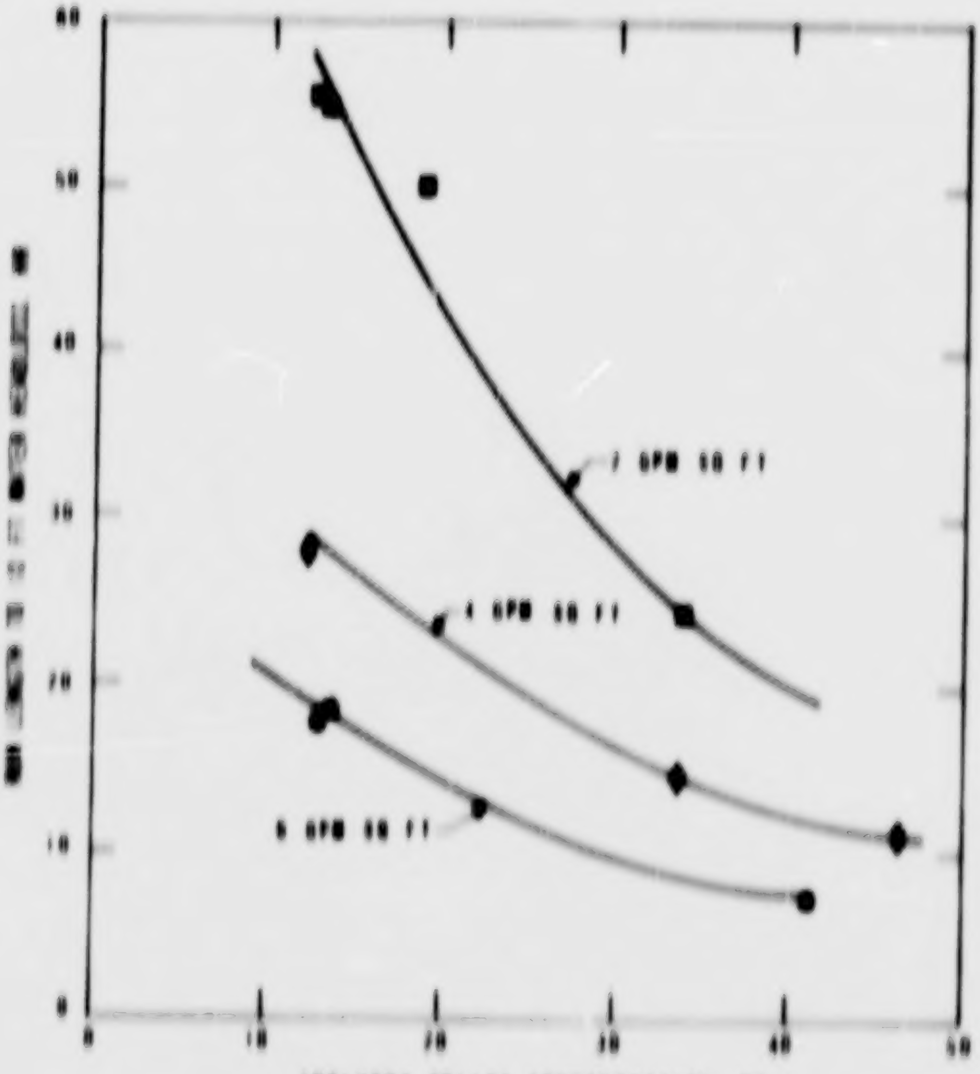


TABLE 7-11
EXPECTED FILTER PERFORMANCE FOR ACTIVATED SLUDGE PLANTS (15)

Type of Activated Sludge Process	Good Biological Treatment				
	Filter Influent		Filter Effluent		
	BOD mg/l	SS mg/l	BOD mg/l	SS mg/l	Run Time hr
Conventional and Extended Aeration	12-15	15-25	2-5	1-4	16-24
Contact Stabilization	15-20	15-25	5-10	1-5	12-20

Type of Activated Sludge Process	Fair Biological Treatment				
	Filter Influent		Filter Effluent		
	BOD mg/l	SS mg/l	BOD mg/l	SS mg/l	Run Time hr
Conventional and Extended Aeration	20-35	30-50	5-10	5-10	6-12
Contact Stabilization	30-45	25-50	20-25	5-10	6-10

TABLE 7-12
EXPECTED FILTER PERFORMANCE FOR TRICKLING FILTER PLANTS (15)

Percent Soluble BOD Removed in Secondary Process

85 Percent					60 Percent				
Filter Influent		Filter Effluent			Filter Influent		Filter Effluent		
BOD mg/l	SS mg/l	BOD mg/l	SS mg/l	Run Time hr	BOD mg/l	SS mg/l	BOD mg/l	SS mg/l	Run Time hr
30-40	30-40	20-30	15-30	6-11	40-50	35-45	30-40	20-25	5-9

TABLE 7-13
IN-DEPTH FILTRATION OF ACTIVATED SLUDGE AND TRICKLING FILTER PLANT EFFLUENTS

Location	Influent Source	Type of Filter	Media	Hydraulic Loading gpm/sq ft	Filter Influent BOD mg/l	Filter Influent SS mg/l	Filter Effluent BOD mg/l	Filter Effluent SS mg/l	Length of Filter Runs hr	Reference
Coldwater, Michigan	Trickling Filter	Horizontal Pressure	Coal [illegible] Sand [illegible] Garnet		[illegible]	[illegible]	[illegible]	[illegible]	[illegible]	[illegible]
[illegible], Michigan	Activated Sludge	Horizontal Pressure	Multimedia		[illegible]	[illegible]	[illegible]	[illegible]	[illegible]	[illegible]
Walled Lake-Novi, Michigan	Activated Sludge	Gravity Downflow	Multimedia	[illegible]		[illegible]		[illegible]		[illegible]
[illegible], Michigan			Multimedia		[illegible]	[illegible]	[illegible]	[illegible]		[illegible]
State College, Pa. (Spring Creek)	Activated Sludge				[illegible]	[illegible]	[illegible]	[illegible]	[illegible]	[illegible]
[illegible], Kentucky ([illegible])	Activated Sludge	Pressure	Multimedia	[illegible]	[illegible]	[illegible]	[illegible]	[illegible]		[illegible]
Ann Arbor, Michigan	Activated Sludge		Multimedia	[illegible]	[illegible]	[illegible]	[illegible]	[illegible]		[illegible]
[illegible], Oregon	Extended Aeration [illegible]	Pressure Downflow	Multimedia	[illegible]	[illegible]	[illegible]	[illegible]	[illegible]	[illegible]	[illegible]
Ventura, California	Trickling Filter	Gravity [illegible] Downflow	[illegible] mm Sand	[illegible]	[illegible]	[illegible]	[illegible]	[illegible]	[illegible]	[illegible]
Hanover Park, Illinois	Activated Sludge	Pressure Upflow	Multimedia	[illegible]	[illegible]	[illegible]	[illegible]	[illegible]		[illegible]
	Activated Sludge	Pressure Upflow	Multimedia	[illegible]	[illegible]	[illegible]	[illegible]	[illegible]		[illegible]
	Activated Sludge	Pressure Upflow	Multimedia	[illegible]	[illegible]	[illegible]	[illegible]	[illegible]		[illegible]
	Activated Sludge		Multimedia	[illegible]	[illegible]	[illegible]	[illegible]	[illegible]		[illegible]
	Activated Sludge		Multimedia	[illegible]	[illegible]	[illegible]	[illegible]	[illegible]		[illegible]

Note: ( ) indicates range of values.

TABLE 7-4
SURFACE FILTRATION OF ACTIVATED SLUDGE AND TRICKLING FILTER PLANT EFFLUENTS

Influent Source	Type of Filter	Media	Hydraulic Loading gpm/sq ft	Filter Influent BOD mg/l	Filter Influent SS mg/l	Filter Effluent BOD mg/l	Filter Effluent SS mg/l	Reference
Activated Sludge	Gravity Downflow	Coal - 30 in. Sand - 12 in. Garnet - 6 in.	2.2	20	16	7	7	22
Activated Sludge	Gravity Downflow	Coal - 30 in. Sand - 12 in. Garnet - 6 in.	4.0	25	15	7	5	22
Activated Sludge	Gravity Downflow	Coal - 30 in. Sand - 12 in. Garnet - 6 in.	2.9	19	12	7	6	22
Activated Sludge	Gravity Downflow	—	1.0-6.0	–	–	32-78%[1]	72-93%[1]	23
Activated Sludge	Gravity Downflow	—	2.0	–	–	57%[1]	46%[1]	24
Activated Sludge	Gravity Downflow	—	2.0-6.0	–	–	80%[1]	70%[1]	25
Contact Stabilization	Gravity Downflow	—	3.3	18	14	4	3	26
Trickling Filter	Gravity Downflow	—	0.75	47	36	17	10	26

[1] Removal Efficiencies.

TABLE 7-15

FILTRATION OF CHEMICALLY TREATED SECONDARY EFFLUENT

[illegible]

While it is recognized that filtration efficiency varies depending on the type of filter, bed design, characteristics of the wastewater, and many other factors previously discussed, it is informative to examine the limits of this variability. An evaluation of filtration data from a number of installations indicates that removal of SS averages approximately 70 percent, with a range of 50 to 90 percent.

7.5 Activated Carbon Adsorption

The limitations of conventional biological treatment processes in regard to reliable achievement of a high degree of organic removal (particularly of certain compounds which are refractory to biodegradation), along with increasingly strict water quality standards, emphasize the need for a supplementary organic removal process. Thus, activated carbon is presently being used to provide tertiary treatment of biologically treated effluents. Experience gained from the operation of activated carbon plants for tertiary treatment of wastewater suggests that activated carbon need not be restricted to a polishing role, but can be used as an alternative to biological treatment. Replacement of conventional biological treatment by activated carbon (i.e. secondary treatment application) is emphasized in the Process Design Manual for Carbon Adsorption (2). The following discussion is concerned exclusively with the tertiary application of carbon.

Activated carbon for wastewater treatment can be used either in the powdered or in the granular form. The impracticality of economical regeneration has restricted the use of powdered carbon in wastewater treatment, although this problem is being resolved.

7.5.1 Process Principles and Design Factors

The adsorption of organic materials from wastewater onto the activated carbon involves complex physical and chemical interactions. Biological degradation of adsorbed materials also occurs, and this can significantly enhance the overall treatment performance (10) (11).

The ability of activated carbon to adsorb large quantities of dissolved materials from wastewater is due to its highly porous structure and to the resulting large surface area, which provides many sites for adsorption of dissolved materials.

Important factors in the design of activated carbon treatment facilities include: pretreatment requirements; particle size; hydraulic loading and contact time; regeneration losses; flow configuration; and required effluent quality.

7.5.1.1 Pretreatment Requirements

Granular carbon can be used as a direct polishing technique for secondary effluents or may be preceded by a SS removal process or other treatment processes as required to accomplish treatment objectives. Long term studies at Pomona, California, have indicated that carbon can function as an effective filter in addition to removing dissolved organic matter (34). However, it is important that the secondary treatment plant clarification system produce a reasonably high quality effluent (<20 mg/l SS), since the use of carbon as a media to remove large quantities of suspended material is not a cost effective treatment technique.

7.5.1.2 Particle Size

Theoretically, carbon particle size primarily affects the rate of adsorption and not the capacity of the carbon. Adsorption rates are greater for smaller particle sizes than for larger particle sizes. However, adsorbents close to saturation will be less affected by particle size than adsorbents in their virgin state (32).

Data from Lake Tahoe indicates that there will be a reduction in the adsorption capacity of about 30 to 35 percent in going from 12 by 40 mesh carbon to 8 by 30 mesh carbon at a relatively short contact time (32). This apparent difference in adsorption capacity attributable to particle size is minimized at longer contact times (35). Since finer particle sizes are susceptible to greater head losses, 12 by 40 mesh carbon is probably not suitable for use in downflow columns (35).

7.5.1.3 Hydraulic Loading Rate and Contact Time

Contact time, hydraulic loadings, and bed depth are interrelated physical parameters. Of the three, contact time is clearly the most important. Since the activated carbon treatment of wastewater requires that a definite contact time be established to complete the adsorption process, any increase in applied hydraulic load necessitates a deeper carbon column to maintain the same contact time.

Data obtained at the Pomona, California Pilot Plant indicate that Total Organic Carbon (TOC) removal does not vary significantly after 15 minutes contact time for hydraulic loading rates of 4, 7, and 10 gpm/sq ft (32). It was further noted that for equivalent contact times, the percent TOC removal was similar for hydraulic loading rates of 4, 7 and 10 gpm/sq ft. These results indicate that contact time is more important than applied hydraulic loadings, and is, in fact, the most important design factor in carbon adsorption systems.

Typical hydraulic loading rates and contact times used in various locations are shown in Table 7-16. It should be noted that both gravity and pressure systems are available. Gravity flow systems are not likely to be practical at hydraulic loading rates above about 4 gpm/sq ft.

TABLE 7-[illegible]

TYPICAL CARBON COLUMN DESIGN DATA

[illegible]	[illegible]	[illegible]	[illegible]	[illegible]	[illegible]	[illegible]	[illegible]	[illegible]	[illegible]
1. [illegible]	[illegible]	[illegible]	[illegible]	[illegible]	[illegible]	[illegible]	8 x 30	[illegible]	[illegible]
2. [illegible]	[illegible]	[illegible]	[illegible]	[illegible]	[illegible]	[illegible]	8 x 30	[illegible]	[illegible]
3. [illegible]	[illegible]	[illegible]	[illegible]	[illegible]	[illegible]	[illegible]	8 x 30	[illegible] [illegible]	[illegible]
4. [illegible]	[illegible]	[illegible]	[illegible]	[illegible]	[illegible]	[illegible]	8 x 30	[illegible]	[illegible]
5. [illegible]	[illegible]	[illegible]	[illegible]	[illegible]	[illegible]	[illegible]	8 x 30	[illegible]	[illegible]
6. [illegible]	[illegible]	[illegible]	[illegible]	[illegible]	[illegible]	[illegible]	8 x 30	[illegible] [illegible]	[illegible]
7. [illegible]	[illegible]	[illegible]	[illegible]	[illegible]	[illegible]	[illegible]	8 x 30	[illegible] [illegible]	[illegible]
8. [illegible]	[illegible]	[illegible]	[illegible]	[illegible]	[illegible]	[illegible]	8 x 30	[illegible]	[illegible]
9. [illegible]	[illegible]	[illegible]	[illegible]	[illegible]	[illegible]	[illegible]	8 x 30	[illegible]	[illegible]
10. [illegible]	[illegible]	[illegible]	[illegible]	[illegible]	[illegible]	[illegible]	8 x 30		
11. [illegible]	[illegible]	[illegible]	[illegible]	[illegible]	[illegible]	[illegible]	8 x 30	[illegible] [illegible]	[illegible]
12. [illegible]	[illegible]	[illegible]	[illegible]	[illegible]	[illegible]	[illegible]	12 x 40	[illegible]	[illegible]
13. [illegible]	[illegible]	[illegible]	[illegible]	[illegible]	[illegible]	[illegible]	12 x 40	[illegible]	[illegible]

[1] [illegible]

[2] [illegible]

7.5.1.4 Effect of Regeneration

Activated carbon requires regeneration when its adsorption capacity is exhausted. Considerable effort has been expended to determine the effect of regeneration on adsorption capacity of the carbon. However, since few research groups have regeneration facilities, only limited data are available. Results obtained at Pomona (32) indicate that the adsorptive capacity decreases by approximately 35 percent after seven regeneration cycles, as indicated on Figure 7.4. It was also determined that regeneration does not affect the degree of organic removal in subsequent exhaustion cycles. This loss of capacity is not necessarily a critical factor, since it is necessary to make up physical losses of carbon after each regeneration cycle. These losses are caused by several factors: carbon is burned and lost through the stack as combustion products; or is abraded into dust in the course of handling. Further "losses" are due to the buildup of inorganic ash in the carbon particles during repeated use and regeneration.

7.5.1.5 Flow Configuration

Depending on the dissolved organic and SS loading, any of several optional flow configurations can be adopted:

1. Downflow Beds in Series — the lead contactor is removed, regenerated, and replaced in line at the downstream end, the other contactors being moved up in sequence.

2. Downflow Beds in Parallel — parallel beds are arranged in a staggered exhaustion pattern so that when one is exhausted and removed from service, the product of the others can be blended with that portion of flow normally treated by the exhausted contractor to maintain the required product quality for the entire plant.

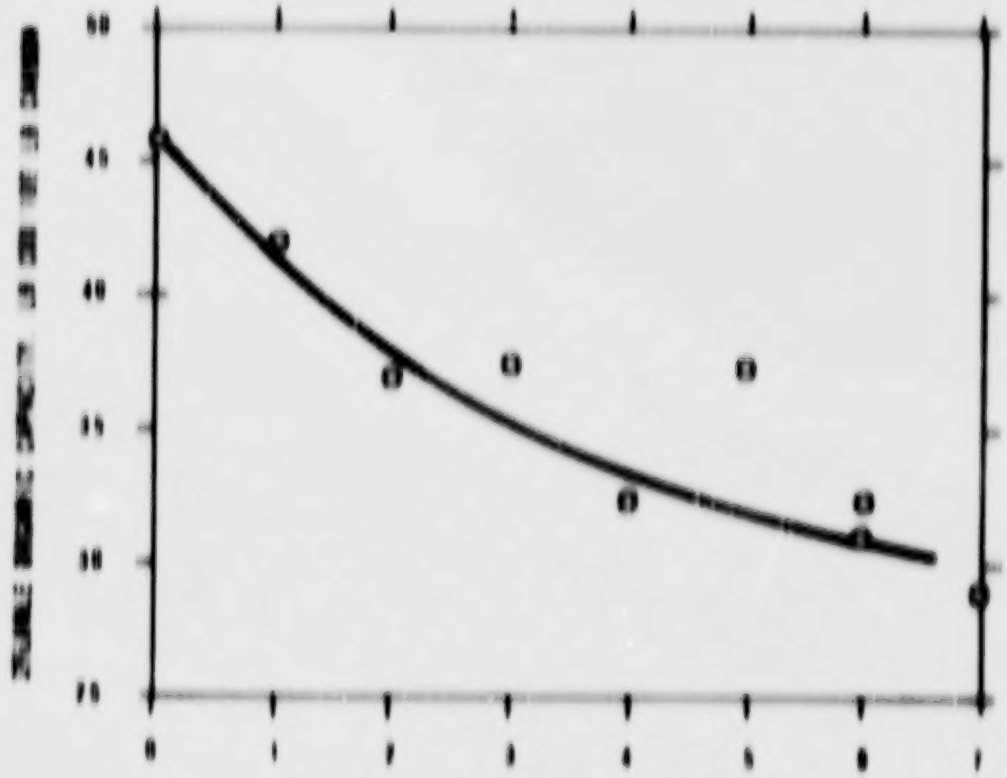
3. Upflow Beds (expanded or partially expanded) — no head loss is built up, and no backwashing is necessary; postfiltration is required; the same series and parallel considerations apply as for downflow operation.

4. Upflow (moving bed) — exhausted lower strata of the bed are continuously removed and replaced at the top of the bed by virgin carbon.

Parallel comparison studies between downflow and upflow beds were conducted at the treatment plant of the Ewing-Lawrence Sewerage Authority located near Trenton, New Jersey (36). These studies have indicated that upflow and downflow beds have equivalent adsorption capacities since each system removed approximately the same quantity of soluble organic material from the plant's secondary effluent (see Table 7.17). The downflow bed system is more effective for removal of SS, but at the expense of an increased head loss.

FIGURE 7-4

EFFECT OF REACTIVATION ON ADSORPTION CAPACITY



and a corresponding increase in the frequency of backwashing. Accordingly, an upflow bed system can process more effluent than a downflow bed system of equivalent size because of less downtime for the carbon backwash operation. If no allowance is made for backwashing, an upflow bed system is capable of processing about 9 percent more effluent than a downflow system of equivalent size (46).

TABLE 7-17
PERFORMANCE OF UPFLOW BED AND
DOWNFLOW BED ADSORBERS (46)

Description	Filtered Secondary Effluent		Unfiltered Secondary Effluent	
	Downflow	Upflow	Downflow	Upflow
TOC Removed, percent	53.6	48.1	57.6	52.8
Soluble Organic Carbon Removal, percent	42.3	44.7	49.6	45.7
Soluble Organic Carbon Removed per lb Active Carbon, lb	0.19	0.20	0.24	0.22

7.5.1.6 Performance

In addition to the above design considerations, the question of effluent quality standards should not be neglected.

The effectiveness of granular activated carbon for upgrading the treatment efficiency is well established. Full-scale operating experiences at Pomona and South Lake Tahoe, California; Colorado Springs, Colorado; and Piscataway, Maryland have left little doubt regarding process efficiency, operating cost and reliability of these systems. The principal design parameters for these plants and other tertiary granular carbon plants were previously shown in Table 7-16.

Performance data for operating tertiary carbon facilities are shown in Table 7-18. It is evident from these data that a very high quality effluent can be produced with the addition of tertiary carbon treatment.

7.5.2 Laboratory and/or Pilot Plant Investigations

Activated carbon removes dissolved materials from wastewaters by a combination of three mechanisms: adsorption, filtration, and biological degradation. Therefore, in order to judge

TABLE 7-[illegible]

PERFORMANCE OF TERTIARY CARBON WASTEWATER TREATMENT PLANTS

	Pilot Plant Pomona, Calif. (37)		Lake Tahoe, Calif. (2)			[illegible] (27)			Pilot Plant Orange County, Calif. (2)	
	Carbon		Secondary	Carbon		Secondary	Carbon		Carbon	
	Influent	Effluent	Effluent	Influent	Effluent	Effluent	Influent	Effluent	Influent	Effluent
BOD, mg/l	3	1	30	3	0.7				[illegible]	2
COD, mg/l	43	20	70	25	10	32.9	12.6	6.1	[illegible]	[illegible]
TOC, mg/l	12	3				11.9	7.3	1.5		
SS, mg/l	4	0.6	25	0	0	15.7	[illegible]	1.0	[illegible]	1
Turbidity, Jtu	8	1	15	0.2	0.3					
Color (Platinum-Cobalt)	30	3								
Odor	12	1								
MBAS, mg/l	–	–	2.9	0.5	0.1				3.4	0.4

the effectiveness of activated carbon for wastewater treatment, both laboratory and pilot testing are required.

The adsorption mechanism can be evaluated in the laboratory by running adsorption isotherms. Actual plant conditions should be simulated with regard to temperature, pH, and pretreatment. A detailed isotherm procedure is given in many books, as well as in the Process Design Manual for Carbon Adsorption (*2*).

Adsorption isotherms are normally conducted by contacting a sample of wastewater with varying amounts of pulverized carbon for a standard interval of time. The wastewater sample is analyzed for TOC, COD, or BOD (as deemed necessary), both before and after contacting with the pulverized activated carbon. The treated water should be coarse filtered prior to analysis to eliminate carbon fines. The isotherm is a plot of the amount of solute adsorbed per unit weight of carbon as a function of residual concentration of solute. The isotherm is empirically represented by the following expression (23):

$$x/m = KC^{1/n}$$

where:

x = weight of solute adsorbed
m = weight of carbon
C = equilibrium concentration of solute in solution after adsorption
K and n are constants

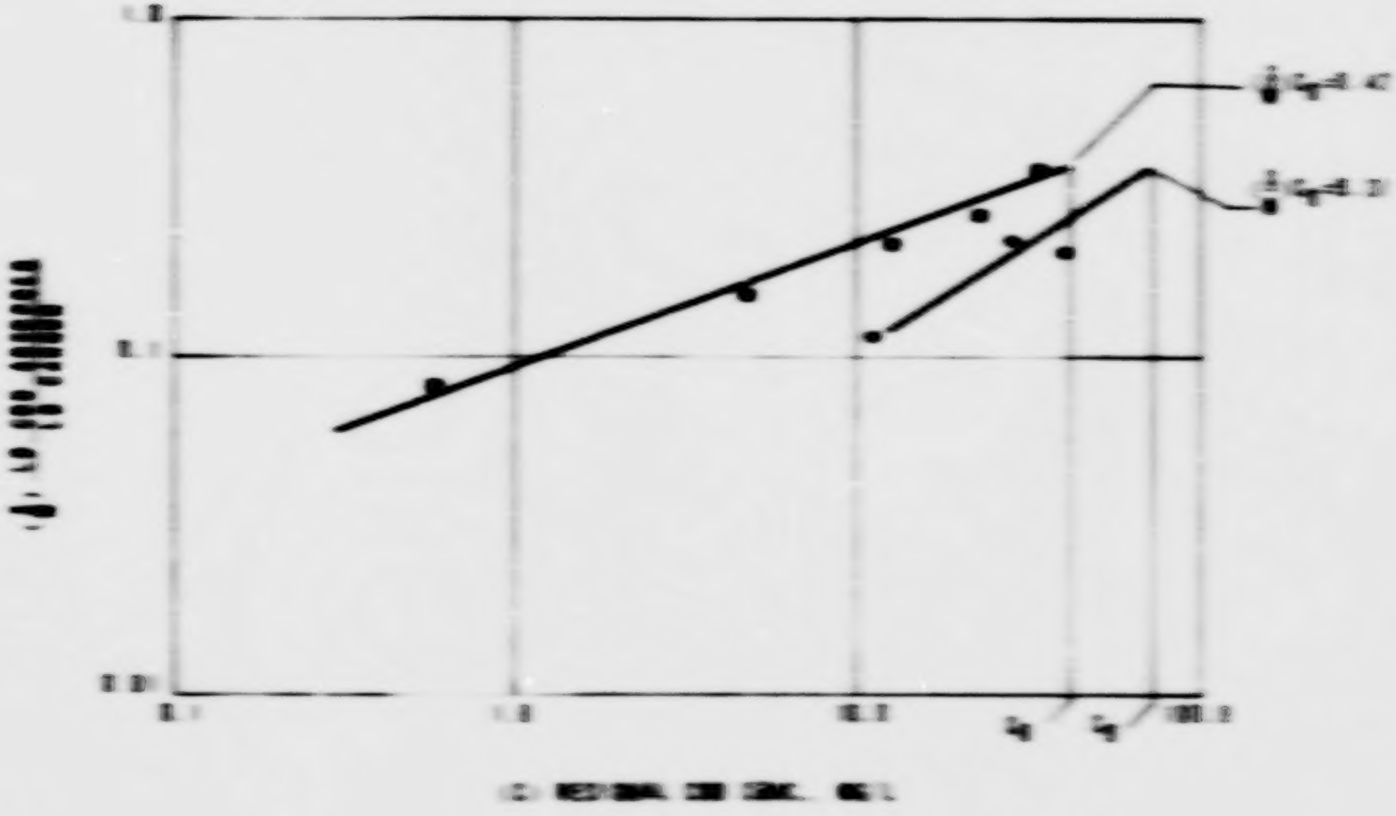
The isotherms are normally plotted on a log-log scale. The extrapolation of the isotherm line to the initial concentration (abscissa) gives the theoretical adsorption capacity of that carbon when it is in equilibrium with the influent concentration.

The advantage of isotherms are: (1) they are relatively simple tests to perform; (2) they indicate whether the desired degree of treatment can be readily achieved; and (3) they give the approximate adsorptive capacity of the carbon in a column application. However, isotherm results should not be used to extrapolate carbon capacities and dosages to full scale plant size.

Typical isotherms obtained for the same carbon with different secondary wastewater effluents were reported by Masse (17) (see Figure 7-5). The results shown on Figure 7-5 indicate that the adsorptive capacity of the carbon with respect to COD varies from 0.17 to 0.12 lb of COD/lb of carbon. This is equivalent to 1.50 to 1.00 lb of carbon/1,000 gal of throughput. The carbon requirements (per unit volume of wastewater treated) obtained from isotherms are conservative (i.e., high), because removal by adsorption alone is estimated.

FIGURE 7-5

COD ISOTHERMS USING VIRGIN CARBON
AND DIFFERENT SECONDARY WASTEWATER EFFLUENTS (27)



Since isotherms cannot measure the quantity of organics removed by filtration and biological action, pilot column testing must be conducted to evaluate the effect of these factors. Column testing helps to determine: (1) the required contact time; (2) the adsorptive capacity of the carbon; (3) the pressure drop across the beds and backwash requirements for downflow operation; and (4) the shape of the column exhaustion wave front.

The column used in pilot testing should have a diameter of at least 4 inches. The depth of column depends on the range of contact times being considered, as does the hydraulic loading. Normally, two to four columns are used in series, since this arrangement permits evaluation of the effect of different contact times on effluent quality. When loading granular carbon into the test column (a "wet" packing procedure is recommended), care must be exercised to avoid entrapping air within the carbon column. Air entrapment causes channeling and reduces the contact area, which in turn yields false test results. The Process Design Manual for Carbon Adsorption (2) describes the conduct of pilot operations in more detail.

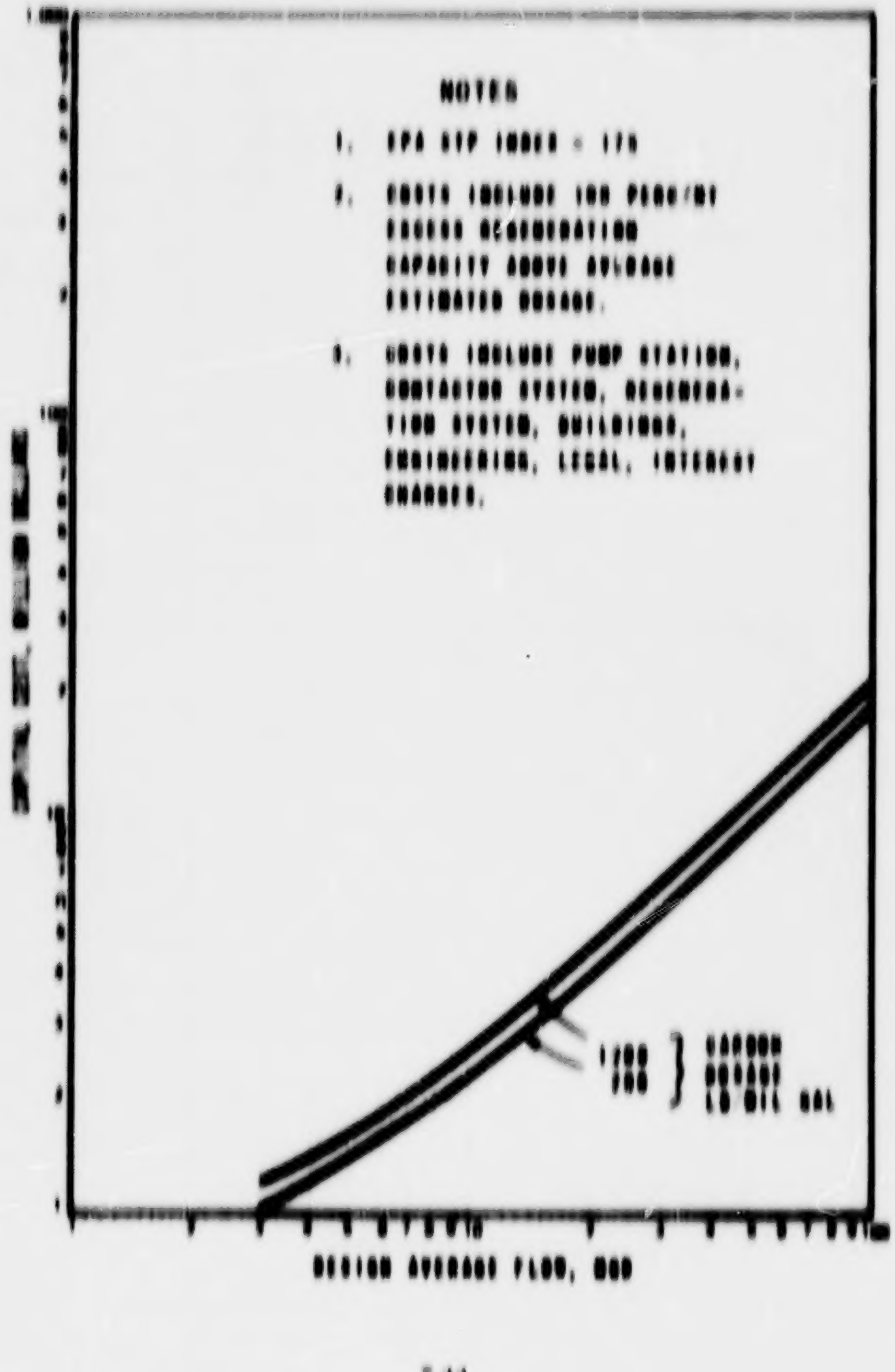
7.5.4 Costs of Tertiary Carbon Treatment

The capital cost of tertiary granular carbon systems will vary widely depending on the particular system design and pretreatment provided. Direct application of secondary effluent to downflow carbon adsorption columns as practiced at Pomona, California, will result in a smaller capital investment than a tertiary system using effluent polishing prior to carbon treatment. Total capital costs for carbon treatment may be estimated from Figure 7-6 for two levels of carbon dosage. For tertiary treatment, the carbon dosage will generally be in the 200 to 500 lb/mil gal range.

Operating cost will depend primarily on the organic loading and associated carbon dosage. Annual costs for power, labor, and maintenance may be estimated from Figure 7-7 which is based on the experience at Lake Tahoe. Total annual and unit costs may be estimated from Figure 7-8 for various carbon dosages.

FIGURE 7-6

TOTAL CAPITAL COSTS FOR CARBON TREATMENT (2)



NOTES

1. EPA STP INDEX = 175
2. COSTS INCLUDE 100 PERCENT EXCESS REGENERATION CAPACITY ABOVE AVERAGE ESTIMATED DOSAGE.
3. COSTS INCLUDE PUMP STATION, CONTACTOR SYSTEM, REGENERATION SYSTEM, BUILDINGS, ENGINEERING, LEGAL, INTEREST CHARGES.

FIGURE 7-7

CARBON ADSORPTION
OPERATION AND MAINTENANCE COSTS (2)

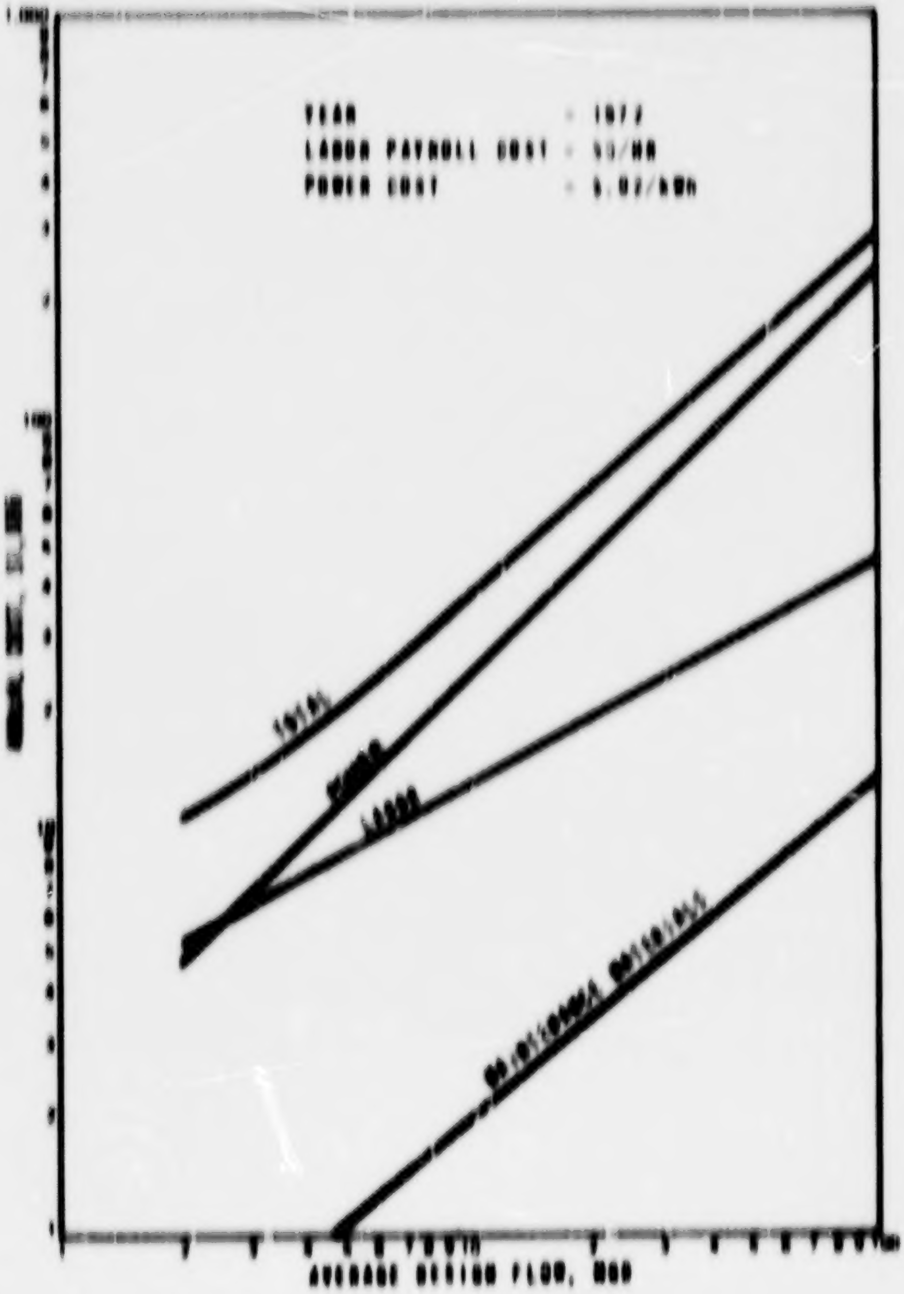
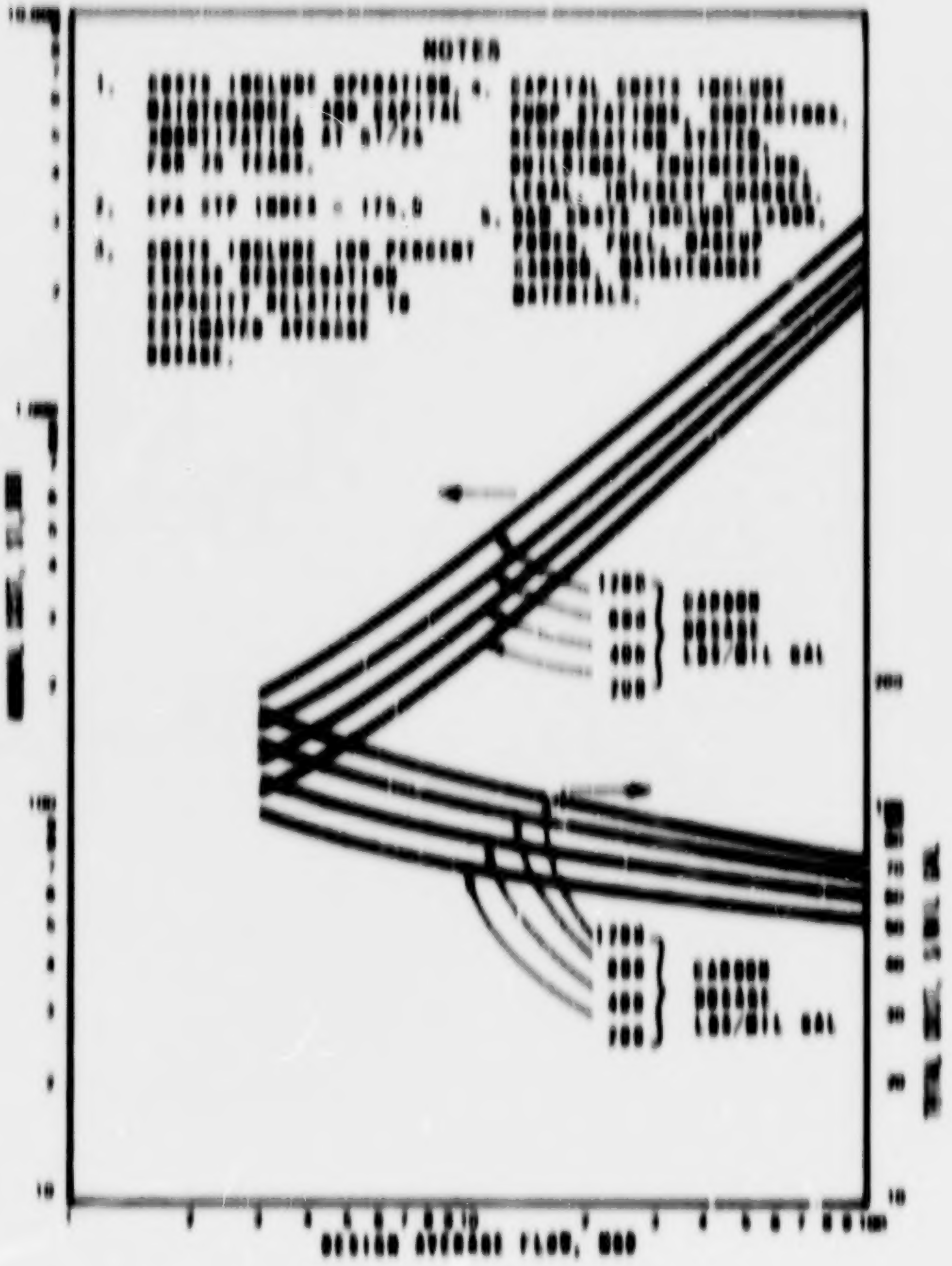


FIGURE 7-8

TOTAL ANNUAL AND UNIT COSTS FOR CARBON TREATMENT (2)



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CHAPTER 8

PREAERATION AND POSTAERATION PRACTICES

8.1 Preaeration

Preaeration of wastewater has been practiced for over 30 years throughout the United States, generally for the purpose of odor control and to improve the treatability of the wastewater. Short aeration periods ranging up to 15 minutes have been found adequate for these purposes. For longer aeration periods, the additional benefits of grease separation and improved flocculation of solids have also been observed (1).

While the use of aerated grit chambers is becoming increasingly popular as a pretreatment unit in wastewater treatment plants, their use should not be expected to substantially increase BOD or SS removal during primary clarification due to the relatively short detention times normally employed.

8.1.1 Preaeration Process Design Considerations

The major parameters to be considered in the design of preaeration facilities are rate of air application and detention time. In order to maintain proper agitation, the air supply system should provide a range of 1.0 to 4.0 cfm per lineal foot of tank. This range will assure adequate performance for nearly all physical tank layouts and types of aeration equipment used.

Effective preaeration has been achieved at detention times of 45 minutes and less (1) (2). The Ten States Standards (3) recommends a detention time of 30 minutes for effective solids flocculation when inorganic chemicals are used in conjunction with preaeration. For appreciable BOD reduction, a minimum of 45 minutes is recommended. In addition, the use of polyelectrolytes may vary these detention times.

8.1.2 Upgrading Considerations for Preaeration

Several efforts have been made to determine the effects of preaeration on primary clarifier performance. In 1961, Seidel and Baumann (2) conducted a study at the Ames, Iowa, secondary treatment plant to determine the effects of preaeration by direct comparison with a parallel primary clarifier receiving an equal flow of the same waste without preaeration. They determined that with 45 minutes detention and an aeration rate of 0.1 cu ft/gal of wastewater, BOD and SS removals were both increased by seven to eight percent in the primary tank. In this instance, an economic evaluation based on conventional design standards showed that preaeration costs increase the total annual plant operating cost by two to three percent.

In an analysis of operation data from 30 plants using preaeration, Roe (1) observed increased primary clarifier SS removals were obtained in some instances. Quantities of air used varied between 0.06 and 0.15 cu ft/gal. However, informal tests at the Allegheny County Sanitary Authority Wastewater Treatment Plant showed that little or no improvement in primary clarifier performance was achieved using preaeration with a detention time of 45 minutes.

These results and others as reported by Eliassen (4) have lead to some disagreement as to the desirability of providing preaeration for increased primary SS or BOD removals. Although the studies by Roe, and by Seidel and Baumann have established that primary clarifier performance will improve with prolonged preaeration time, it is often difficult to justify the additional cost since the incremental removals achieved as a result of preaeration probably would be partially or completely obtained in the secondary system without preaeration.

The specific conditions involved in determining the merits of preaeration must be evaluated for each treatment situation. The costs of providing sufficient preaeration and clarifier detention times required for significant improvements in primary performance must be weighed against the savings that would be derived from the resulting reduction in secondary loading.

The most likely upgrading situations involving preaeration that will be encountered are: (1) when an existing plant with preaeration is to be upgraded, and (2) when considering chemical addition prior to existing primary treatment. In the latter case, existing plant hydraulics will have a significant impact on the designer's options.

Where existing preaeration-primary plants are being upgraded to secondary treatment, it may be beneficial to inject waste-activated sludge upstream of preaeration to improve primary clarifier performance.

8.1.3 Preaeration with Chemical Addition

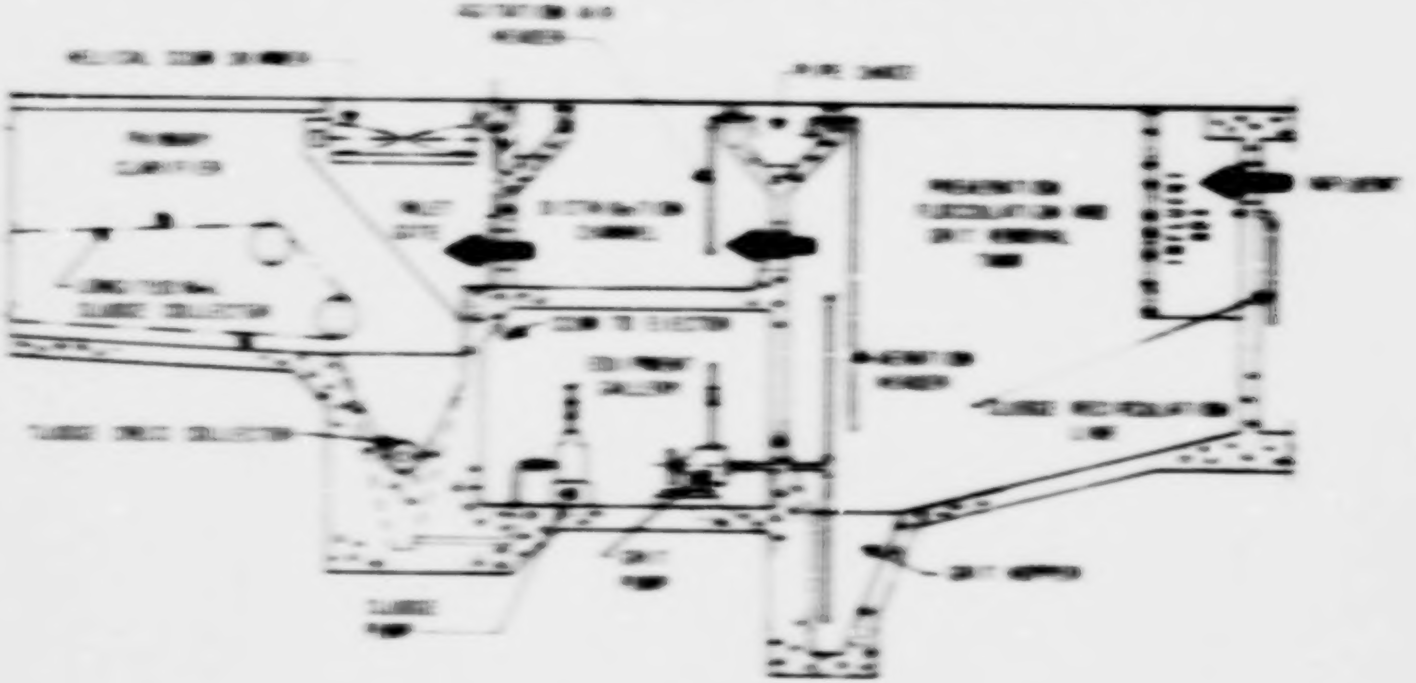
Removal of BOD, SS and phosphorus in primary clarification may be substantially improved through the use of chemical additions to preaeration units (5). The addition of chemicals such as lime, alum, ferric chloride, or organic polyelectrolytes to the wastewater flow ahead of the preaeration unit allows for adequate mixing of the chemical by air stirring. Chemical addition can also reduce the preaeration time required to obtain improved primary clarifier performance. Before implementing preaeration with chemical addition, it is recommended that full- or pilot-scale tests be conducted to determine optimum chemical dosages and removal efficiencies.

The Central Contra Costa Sanitary District near San Francisco, California, has operated a 3 mgd advanced treatment test facility since November 1971. The existing preaeration and primary clarification facilities shown on Figure 8-1 were upgraded by providing lime

FIGURE 8-1

PRIMARY TREATMENT UNITS AT THE

CENTRAL CONTRA COSTA SANITARY DISTRICT WATER RECLAMATION PLANT


AGITATION AIR
HEADER
HELICAL SCUM SKIMMER
PRIMARY
CLARIFIER
INLET
GATE
DISTRIBUTION
CHANNEL
PREAERATION
FLOCCULATION AND
GRIT REMOVAL
TANK
INFLUENT
LONGITUDINAL
SLUDGE COLLECTOR
EQUIPMENT
GALLERY
AERATION
HEADER
SLUDGE RECIRCULATION
LINE
SLUDGE CROSS COLLECTOR
GRIT
PUMP
SLUDGE
PUMP
GRIT HOPPER

addition to precipitate phosphorus and to maximize the removal of raw wastewater solids thus reducing the load on the subsequent combined oxidation-nitrification unit (5) (6).

At the Contra Costa facility, lime is added to the raw wastewater in a deep channel immediately after screening. A hydraulic jump occurs ahead of the preaeration tank and provides the necessary agitation for chemical mixing. Parker and Stites (5) indicate that the combined preaeration and grit removal tank performs well as a flocculation basin using lime as the coagulant. Diffused air mixing is used to promote the formation of large readily settleable flocculent particles. To further enhance coagulation and floc aggregation, settled primary solids are recirculated to the preaeration tank influent. Increasing the solids concentration and solids contact in this manner has been shown to improve flocculation efficiency.

Air diffusers of the swing type were selected for this application as being most compatible with the multiple function served by the preaeration tank. Diffused air induced currents provide a readily adjustable stirring action necessary to promote flocculation and grit collection. Problems with scaling have often been encountered where lime addition is practiced and rag fouling of mechanical pretreatment equipment is common. The easy accessibility of air diffusers for maintenance minimizes these problems.

Critical design criteria for flocculation are detention time and aeration rate. At Contra Costa, the preaeration tank was designed for 30 minutes detention time at average dry weather flow. Satisfactory results have been obtained at operational detention times of 20 minutes. Air supply rates for lime flocculation must be lower than the rates used for conventional preaeration to avoid shearing of the floc. Although optimal rates have not been definitely established, preliminary results based on the Contra Costa experience indicate air requirements may be 1/4 to 1/7 those normally used for preaeration.

At an overflow rate of 1,000 gpd/sq ft and 1.5 hours detention, the preaerated, chemically treated system significantly outperforms a parallel conventional system, as shown in Table 8-1 (6).

8.3 Postaeration

Many states are considering or have already enacted legislation requiring the maintenance of minimum DO concentrations in wastewater treatment plant effluents. This design criteria is established to provide additional oxygen to the receiving waters. Plant effluents from secondary clarifiers normally contain between 0.5 to 2.0 mg/l of DO. Most surface water quality standards, depending on the intended water use, specify a minimum DO concentration of 4.0 mg/l.

TABLE 8-1
PERFORMANCE OF PRIMARY FACILITIES AT CENTRAL CONTRA COSTA SANITARY DISTRICT WITH CHEMICAL TREATMENT TO PREAERATION

	pH 11.5 Operation Ca(OH)$_2$ = 300 mg/l		pH 11.0 Operation Ca(OH)$_2$ = 400 mg/l	
	Control Primary	Chemical Primary	Control Primary	Chemical Primary
BOD Removal, Percent	46	74	37	69
SS Removal, Percent	74	79	71	76
Grease Removal, Percent	44	79	21	64

8.3.1 Postaeration Process and Design Considerations

There are at least four methods available for the postaeration of a wastewater treatment plant effluent. These are shown on Figure 8.2. Most of these devices were initially developed for water treatment and are now being used in the wastewater treatment field.

8.3.1.1 Diffused Aeration

Diffused air aerators are usually placed in concrete tanks which are commonly 9 to 15 feet deep and 10 to 30 feet wide. Ratios of width to depth of 1.5 are most desirable and should not exceed 2.0 if effective mixing is to be obtained. Tank length is governed by the desired detention period, which usually varies from 10 to 30 minutes.

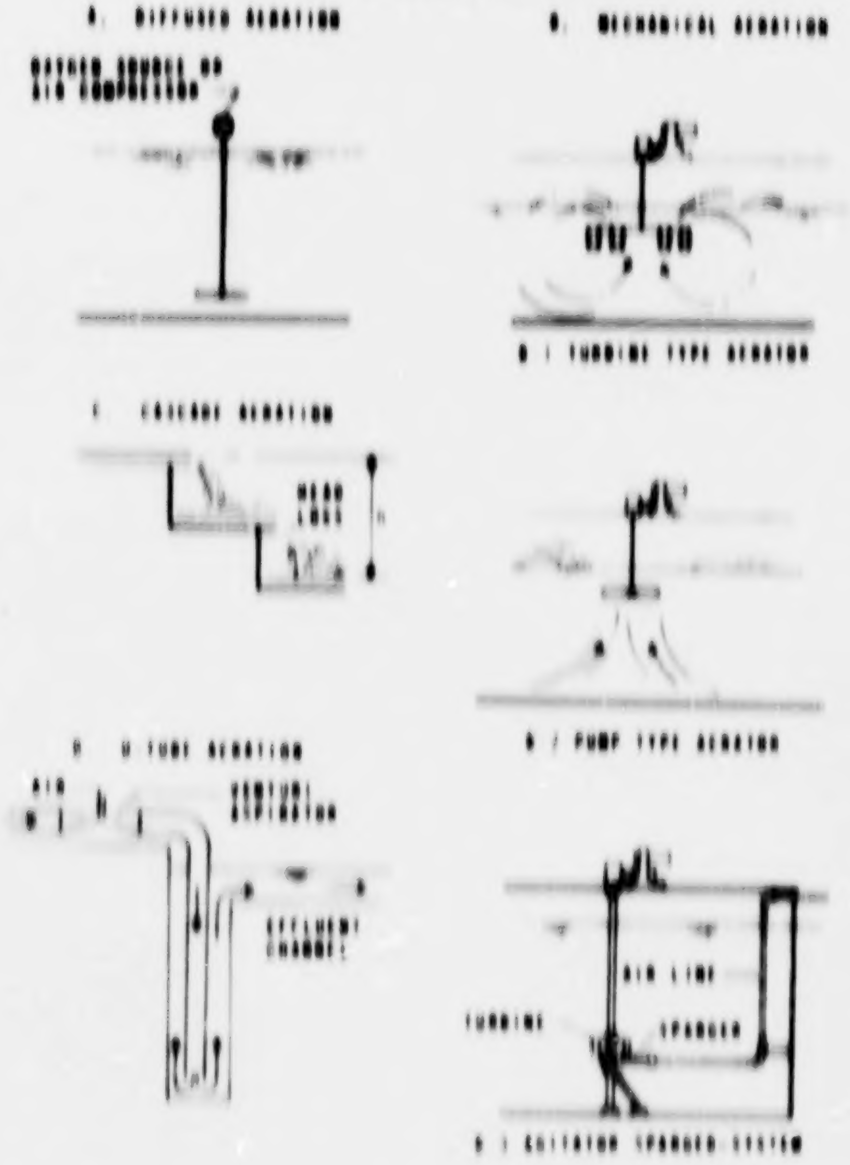
Aeration systems are designed on the basis of their oxygen transfer rate at standard conditions. Standard conditions are defined as 1.0 atmosphere dry pressure at 20 deg C for tap water containing 0.0 mg/l DO. The required rate at which oxygen must be transferred to the wastewater to raise the DO the desired amount under actual operating conditions is given by the following:

$$AOR = 8.34 Q (C - C_0)$$

where:

AOR = Actual oxygen transfer rate, lb O$_2$/day
Q = Postaeration influent flow, mgd
C = Required final DO level after postaeration, mg/l
C$_0$ = DO concentration of the postaeration influent, mg/l

FIGURE 8.2
VARIOUS POSTAERATION DEVICES


A. DIFFUSED AERATION
B. MECHANICAL AERATION
OXYGEN SOURCE OR AIR COMPRESSOR
B-1 TURBINE TYPE AERATOR
C. CASCADE AERATION
HEAD LOSS
h
B-2 PUMP TYPE AERATOR
D. U-TUBE AERATION
AIR
VENTURI ASPIRATOR
EFFLUENT CHANNEL
AIR LINE
TURBINE
SPARGER
B-3 AGITATOR SPARGER SYSTEM

This actual oxygen transfer rate may be adjusted to standard conditions by applying correction factors according to the following equation (7):

$$SOR = \frac{AOR}{\left(\frac{\beta C_s - C}{C_{20}}\right) 1.024^{T-20}\,\alpha}$$

where:

SOR = Standard oxygen transfer rate, lb O_2/day
C_s = DO saturation concentration of tap water at temperature T, mg/l
C_{20} = DO saturation concentration of tap water at 20 deg C, mg/l
T = Design temperature of the wastewater, deg C
α = O_2 transfer coefficient of the wastewater
β = O_2 saturation coefficient of the wastewater

Diffused air systems are designed to provide firm blower capacity, which is the capacity remaining with the largest blower out of service. The maximum air rate required to provide firm capacity may be computed from the standard oxygen transfer rate as follows:

$$A_{fm} = \frac{SOR}{1440 \; e_t \; \gamma_a \; P_O}$$

where:

A_{fm} = Firm blower capacity, cfm
SOR = Standard oxygen transfer rate, lb O_2/day
e_t = O_2 transfer efficiency of O_2 in diffused air
γ_a = Specific weight of air at design temperature and relative humidity, pcf
P_O = O_2 content of dry air, proportion by weight

The use of oxygen aeration in the activated sludge process may eliminate the need for postaeration. Oxygen-aerated mixed liquor discharged to the secondary clarifier usually has a DO of at least 6.0 mg/l (8). In some instances, postaeration with high purity oxygen may be advantageous.

The Reno-Sparks, Nevada Joint Wastewater Treatment Facility provides activated sludge secondary treatment for an average design flow of 20 mgd. State standards requiring that a minimum plant effluent DO concentration of 6.0 mg/l be maintained are met by diffused air postaeration of the secondary clarifier effluent prior to chlorine disinfection and discharge.

over a final effluent weir (9). The postaeration unit consists of a single 210-foot by 60-foot by 15-foot tank that provides a detention time of 60 minutes at average design flow. The diffused air system serves both secondary and postaeration requirements using two blowers with a total capacity range of 7,040 to 16,000 cfm of standard air. The system is regulated on the basis of the secondary aeration air requirements with surplus air used for postaeration.

Examination of plant operating records for the year 1978 showed that an average of 17.0 mgd was treated. Monthly average secondary clarifier effluent DO concentration ranged from 0.3 mg/l to 0.6 mg/l, while the plant discharge DO concentration was observed to average between 6.1 mg/l and 9.4 mg/l.

A recent study was conducted to determine on a daily basis the increase in the DO concentration of the treated effluent due to the postaeration facility alone. Typical data taken during this study are as follows:

Flow Rate, Q : 16.21 mgd
Wastewater temperature, T : 20° C
Postaeration influent DO, C_0 : 0.3 mg/l
Air supply rate, A_m : 6,030 cfm
Observed postaeration DO, C : 7.6 mg/l

Assuming a diffuser oxygen transfer efficiency, e_t, of 10 percent, and oxygen transfer and oxygen saturation coefficients, α and β, of 0.85 and 0.95, respectively, the expected postaeration effluent DO concentration, C, may be computed using the equations stated previously. The computed value for this data is 7.4 mg/l, which is in close agreement with the observed DO concentration of 7.6 mg/l.

8.2.1.3 Mechanical Aeration

Mechanical aerators are generally grouped in two broad categories: turbine types and pump types, as shown on Figure 8.3. In all types, oxygen transfer occurs through a vortexing action and/or from the interfacial exposure of large volumes of liquid sprayed over the surface.

To optimize aeration and mixing and to avoid interference between units, aerator manufacturers have developed criteria for minimum areas and depths, depending on the horsepower of the aerator and the configuration of the impeller.

Mechanical aeration systems are designed on the basis of the horsepower required to produce the needed standard oxygen transfer rate (SOTR), which was defined in the previous section. One aeration design equation proposed by Kormanik for a postaeration basin is (10):

$$P = \frac{SOTR}{24 \times N_0 \times F_g \times \eta}$$

where:

P = Horsepower required
$SOTR$ = Standard oxygen transfer rate, lb O_2/day
N_0 = O_2 transfer efficiency under standard conditions in tap water, lb O_2/hp-hr
F_g = Correction factor related to basin geometry
η = Aerator efficiency correction

8.3.1.3 Cascade Aeration

Cascade aeration takes advantage of the effluent discharge to create a series of steps or weirs over which the flow moves in fairly thin layers, as shown on Figure 8-3. The objective is the maximization of turbulence to increase oxygen transfer. Head requirements vary from three to ten feet, depending upon the initial DO and the desired increase. If the necessary head is not available, effluent pumping is required.

In England, the Water Research Laboratory has performed investigations to quantify as much as possible the layout of cascade aeration schemes. Barrett and others proposed the following formulas (11):

$$r = (C_s - C_0)/(C_s - C)$$

$$h = \frac{r - 1}{0.11\,ab\,(1 + 0.046T)}$$

where:

r = The deficit ratio
C_s = DO saturation concentration of the wastewater at temperature T, mg/l
C_0 = DO concentration of the postaeration influent, mg/l
C = Required final DO level after postaeration, mg/l
a = Water quality parameter equal to 0.8 for a wastewater treatment plant effluent
b = Weir geometry parameter equal to unity for a free weir and 1.3 for the step weirs used in their experimental work
T = Water temperature in deg C
h = Height in feet through which water falls

For example, to raise the DO concentration of a wastewater treatment plant effluent from 0.5 mg/l to 4.5 mg/l at 20 deg C, the computed overall height requirement for a series of step weirs would be approximately 4.0 feet. However, it should be pointed out that the values of the parameters a and b are somewhat arbitrary and need further refinement to substantiate preliminary results.

The wastewater treatment plant at Pittsfield, Massachusetts, a trickling filter plant, is followed by two cascades in parallel, each receiving half of the effluent flow. Each cascade consists of a series of 18 three-foot wide concrete steps with a total horizontal length of 64.5 feet and a total vertical drop of 10.5 feet, as shown on Figure 8-3.

Recently, a study was conducted to determine the increase in the effluent DO concentration resulting from flow over the cascades (12). During the study period the total treated flow averaged 7.0 mgd at 15 deg C. The secondary clarifier effluent DO concentration was observed to vary from 3.9 to 4.2 mg/l, while the aerated final effluent DO varied between 5.8 and 6.2 mg/l. The values computed using Barrett's formula with a = 0.8 and b = 1.1 were six to ten percent higher than the observed final effluent DO concentrations for the corresponding clarifier effluent and temperature. This value of b is reasonable in this instance since the concrete steps involved would be expected to produce more turbulence than a free weir (b = 1.0), but less turbulence than step weirs (b = 1.3).

8.2.1.4 U-Tube Aeration

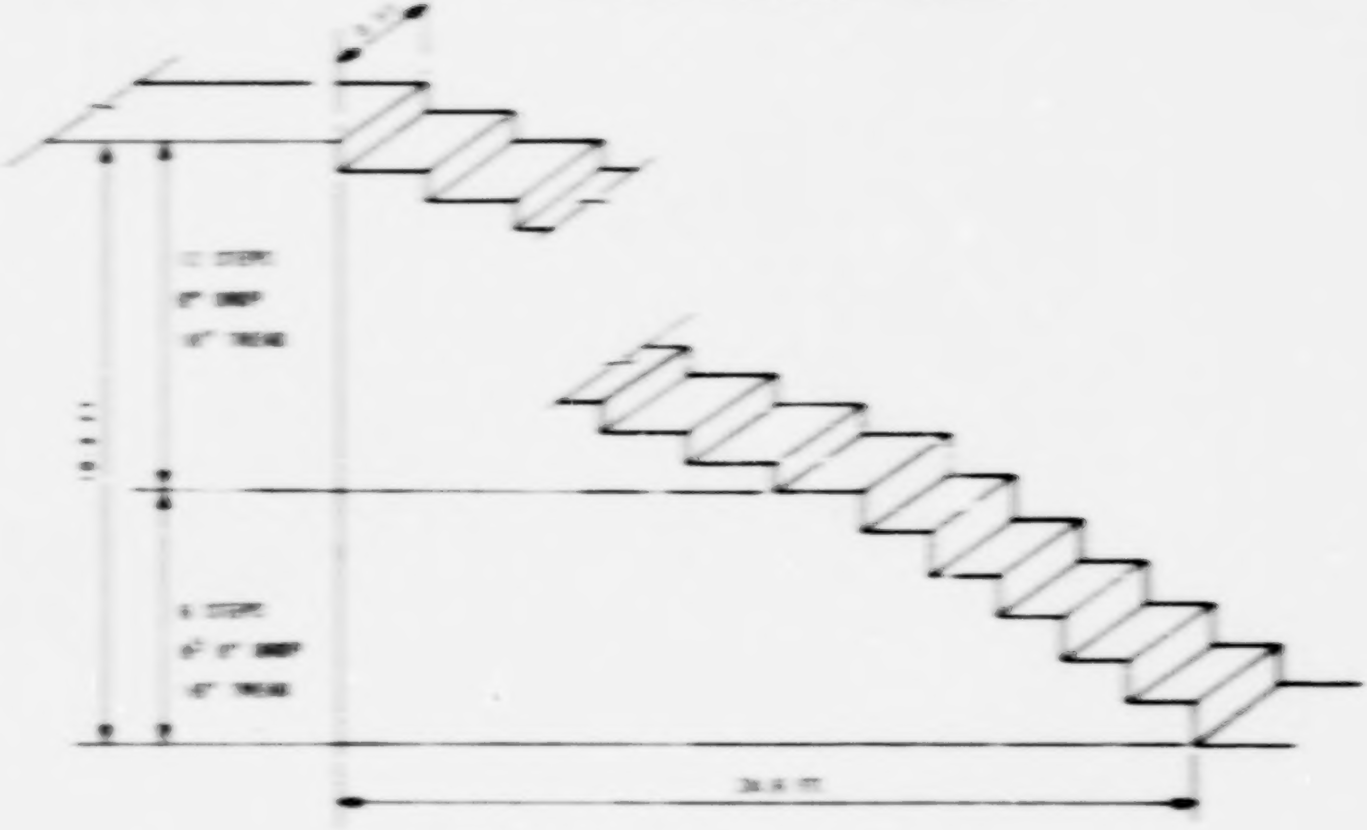
The U-tube aerator consists of two basic components: a conduit to provide a vertical U-shaped flow path and a device for entraining air into the stream flow in the down leg of the conduit as indicated on Figure 8-2. The entrainment device is one of two types: (1) aspirator; or (2) compressor and diffuser. In either case, the entrained air is carried along the down leg of the tube because the water velocity exceeds the buoyant rising velocity of the air bubbles.

Various design considerations include air-to-water ratio, tube cross-sectional area, and depth. The maximum air-to-water ratio practicable is a function of the velocity through the system. At velocities of approximately 4 fps, 20 percent air-to-water injection is about the limit for satisfactory operation (13). The hydraulic head requirements for plants of 5 mgd or less should be less than five feet. If sufficient head is not available, the flow may be pumped through the U-tube.

Speece and Orosco (13) have suggested that one economic method of construction for deep U-tubes, greater than 20 feet in depth, would be a circular hole bored into the soil. The hole would be cased and a smaller pipe then suspended a few feet from the bottom of the hole as shown on Figure 8-2. The diameter of the smaller pipe is selected so that its cross-sectional area is approximately equal to the cross-sectional area of the annular space between the two pipes. Thus, the velocity of the water will be approximately equal in both legs of the U-tube.

FIGURE 8-3

CASCADE AERATOR AT [illegible]


STEPS
DROP
TREAD
STEPS
DROP
TREAD

Presently there are no known U-tube installations in wastewater treatment plants for postaeration. However, full-scale (8 to 20-inch diameter) U-tube units are now being operated to provide in situ aeration in sanitary force mains in two communities in Louisiana and Texas (14). These installations have been effective in reducing serious odor and corrosion problems resulting from sulfides. No maintenance has been required in approximately two years of continuous operation. The possibilities of using a U-tube as a postaeration device seem good at this time.

8.2.3 Upgrading Considerations for Postaeration

The selection of postaeration facilities for treatment plant upgrading is greatly influenced by the specific conditions encountered. The degree to which existing facilities can be used in the upgrading will, in many cases, determine the proper selection of equipment. For example, where the plant is undergoing general upgrading, it may be possible to substantially reduce the costs of mechanical or diffused air facilities by converting an existing basin for use as a postaeration tank.

In some situations there may be operational and/or other advantages in the use of the same type of aeration system for postaeration as is already in use elsewhere in the plant. This is especially true of a diffused air secondary treatment plant where the same blowers and housing structure could be used to supply the postaeration unit as well, with only minor modification.

Cascade postaeration has the obvious advantage of no power, labor or maintenance costs where sufficient head is available between the plant discharge and the receiving water. Where stream standards rather than effluent standards are applicable, but available head is limited at high river levels, it may be possible to use cascade postaeration by taking advantage of seasonal fluctuations in the dilution capacity of the stream. The plant effluent DO is most critical to stream quality during low flow periods. This coincides with the time when the maximum head is available for effective cascade postaeration of the effluent. At higher stream flows when the cascade may be partially submerged, less postaeration is required because of greater dilution.

Although U-tube aeration has been utilized only for sanitary force main aeration to date, the use of this technique for postaeration is promising where sufficient head is available for gravity flow. U-tube postaeration appears particularly attractive where limited space is available for facilities.

Because the peculiarities of each upgrading situation so greatly affect the feasibility and costs of the various alternative postaeration devices, no clear picture emerges as to the relative merits and costs of each method. In each upgrading, an analysis should be made of the applicable conditions such as type of treatment involved, usable existing facilities, and available tank volume and head. As will be discussed in Chapter 9, it may be feasible to combine postaeration and chlorine disinfection advantageously.

It must be noted that postaeration can be expected to produce a detergent type foam. Thus, the facilities should be designed to retain and destroy the foam created, rather than to allow it to flow unimpeded to the surface of the receiving water.

8.3 References

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CHAPTER 9

DISINFECTION AND ODOR CONTROL

9.1 General

Disinfection and odor control are receiving increased attention from regulatory agencies through the establishment and enforcement of rigid bacteriological effluent standards and air pollution standards. In upgrading situations, the need for improved disinfection and the elimination of odor problems are frequently encountered. Adequate and reliable disinfection and odor control are essential to ensure that wastewater treatment plants are environmentally safe and aesthetically acceptable to the public.

9.2 Disinfection

In 1973, 88 percent of all municipal wastewater treatment plants in the United States were equipped with chlorination facilities for disinfection purposes (1). Other disinfectants, such as ozone, ultraviolet light and bromine chloride, are currently being evaluated by the U. S. EPA to determine their potential use in disinfecting wastewater treatment plant effluents. Recent shortages and price escalation of liquid chlorine have emphasized the need to consider other methods of disinfection.

9.2.1 Chlorine Fundamentals

When chlorine is added to water, it hydrolyzes to form hypochlorous acid (HOCl) as follows:

$$Cl_2 + H_2O \rightleftharpoons HOCl + H^+ + Cl^-$$

The hypochlorous acid further ionizes to form hypochlorite ion (OCl⁻) as follows:

$$HOCl \rightleftharpoons H^+ + OCl^-$$

The equilibrium concentrations of HOCl and OCl^- in solution depend on the pH of the wastewater. Increasing the pH will shift the equilibrium toward the formation of higher concentrations of OCl^- and, conversely, decreasing the pH will shift the equilibrium toward the formation of higher concentrations of HOCl.

Chlorine may also be applied as sodium hypochlorite (NaOCl) or calcium hypochlorite [$Ca(OCl)_2$]. The hypochlorite form reacts as follows:

$$NaOCl \longrightarrow Na^+ + OCl^-$$

$$Ca(OCl)_2 \longrightarrow Ca^{2+} + 2OCl^-$$

$$H^+ + OCl^- \rightleftharpoons HOCl$$

The quantity of HOCl plus OCl⁻ in wastewater is known as free available chlorine.

Hypochlorous acid is an extremely potent germicide at pH 6.5 to 7.5 (2). It is also a very active oxidizing agent and is therefore short lived in the presence of readily oxidized compounds such as ammonia. Since most wastewater effluents contain ammonia, the following reactions will occur upon the addition of chlorine:

$$NH_3 + HOCl \longrightarrow NH_2Cl + H_2O$$
(monochloramine)

$$NH_2Cl + HOCl \longrightarrow NHCl_2 + H_2O$$
(dichloramine)

$$NHCl_2 + HOCl \longrightarrow NCl_3 + H_2O$$
(nitrogen trichloride)

The two species that predominate in most cases are monochloramine and dichloramine. They are commonly referred to as the combined available chlorine. The reaction with ammonia is unfortunate, because chloramines are many times less potent than hypochlorous acid as a disinfectant. Figure 9.1 illustrates the relative potency of free and combined forms of chlorine.

Hypochlorous acid also reacts with other organic matter in wastewater such as amino acids and inorganic matter such as sulfides and nitrites to produce chlorine compounds that have very little or no disinfecting power (2). Design engineers should be aware of the extent of such side reactions when determining the optimum chlorine dosages to apply to a wastewater.

Total chlorine residual is the sum of the combined and free chlorine concentrations remaining after a specified period of time. Chlorine demand is defined as the difference between the chlorine applied and the total residual chlorine remaining at the end of the contact period (3). It provides a measure of all chlorine demanding reactions including disinfection.

Browning and McLaren (3) analyzed 833 samples from 12 primary treatment plants and 777 samples from 15 secondary treatment plants to obtain a correlation between residual chlorine and coliform MPN per 100 ml. The results of their analyses are shown on Figure 9.2. The curves represent the MPN remaining after 30 minutes of chlorine contact in

FIGURE 9.1

RELATIONSHIP BETWEEN CONCENTRATION AND TIME FOR 99 PERCENT DESTRUCTION OF ESCHERICHIA COLI BY DIFFERENT FORMS OF CHLORINE AT 2 TO 6°C (8)

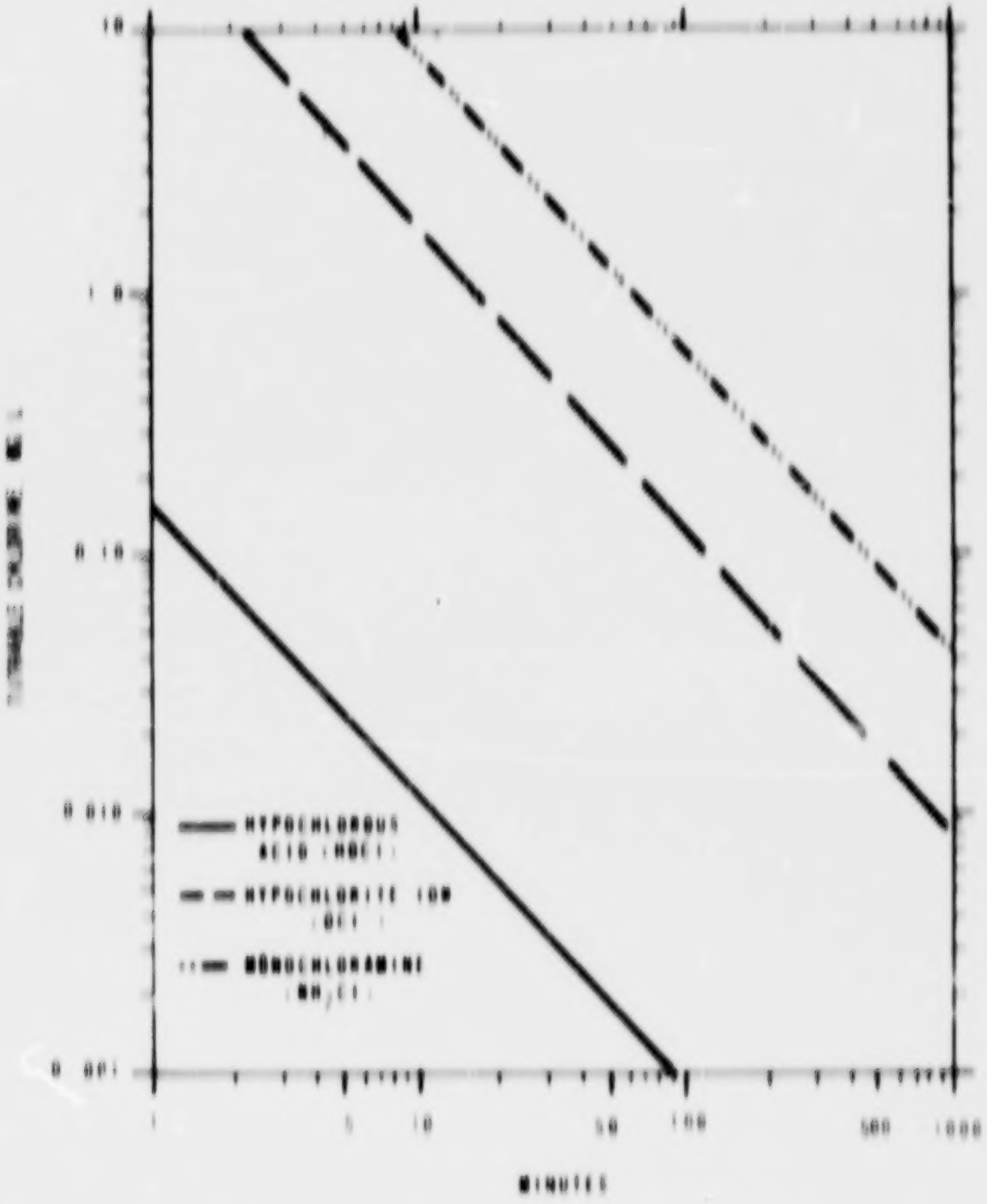
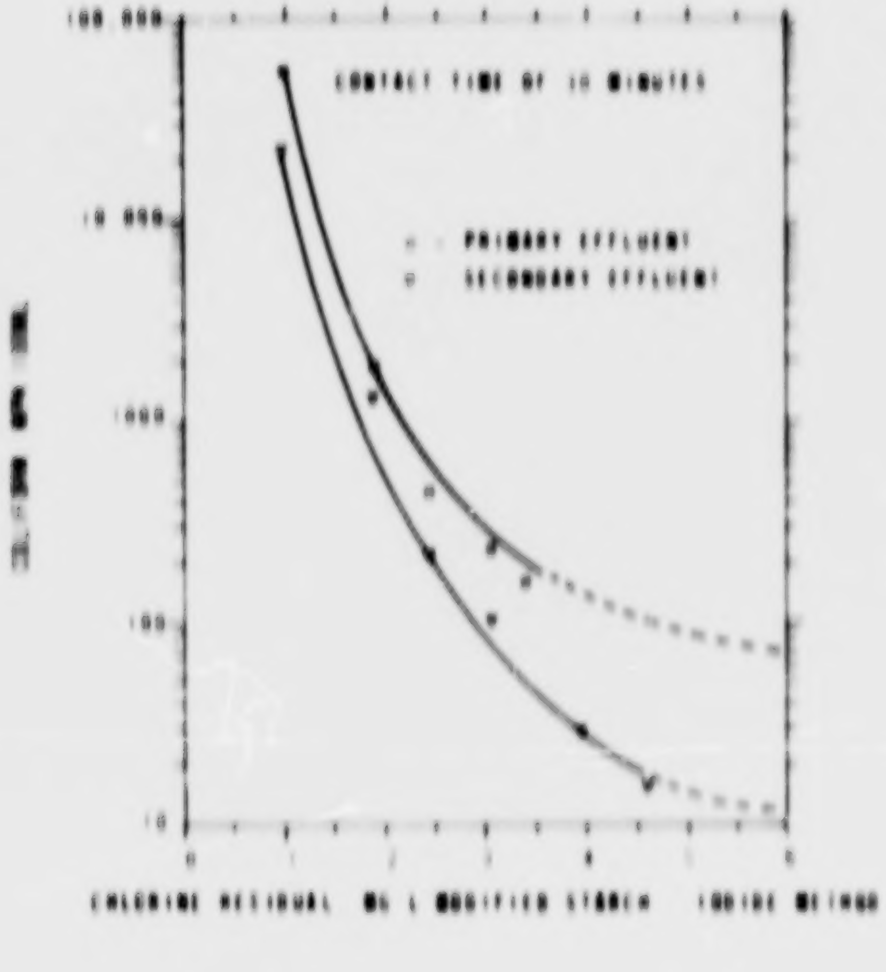


FIGURE 9.2

MPN COLIFORM VS. CHLORINE RESIDUAL (3)



a well designed chlorine contact chamber. Insufficient data were available for residuals above 4 mg/l. The results of these analyses should not be quantitatively applied to any primary or secondary effluent. However, the study provides a good indication of the effectiveness of increased chlorine residuals on primary and secondary effluents.

Other factors which influence the germicidal effectiveness of chlorine are pH and wastewater temperature. At pH values greater than 7.5, the less potent form of free chlorine (OCl^-) predominates. Increased pH also diminishes the disinfecting efficiency of monochloramine (2). It has also been demonstrated that the germicidal effectiveness of free and combined chlorine is markedly diminished with decreasing wastewater temperature (2).

9.2.2 Chlorine Dosage Rates

The following table was taken from the Water Pollution Control Federation's Sewage Treatment Plant Design Manual of Practice No. 8 and contains ranges of chlorine dosages recommended for disinfection (1).

TABLE 9-1
CHLORINE DOSAGE RANGES

Waste	Chlorine Dosage mg/l
Raw Sewage	6 to 12
Raw Sewage (Septic)	12 to 25
Settled Sewage	5 to 10
Settled Sewage (Septic)	12 to 40
Chemical Precipitation Effluent	3 to 10
Trickling Filter Effluent	3 to 10
Activated Sludge Effluent	2 to 8
Sand Filter Effluent	1 to 5

9.2.3 Upgrading Chlorine Contact Tanks

Improved disinfection can be achieved in existing chlorine contact tanks by:

1. Improving the mixing characteristics of the basin.

2. Improving the flow pattern by elimination of short circuiting.

3. Improving the chlorine diffuser system.

4. Lengthening the contact time

5. Increasing the chlorine dosage rate

6. Maintaining the optimum pH range

7. Upstream removal of ammonia nitrogen

The technique to be employed will depend primarily on the disinfection standard to be met, the characteristics of the plant effluent and physical characteristics of the plant itself.

Stringent disinfection standards required for protection of shellfish waters, recreational waters and for various reuse purposes will often require nitrification of the ammonia nitrogen present prior to chlorine addition unless extremely long postchlorination retention basins are used to allow monitoring of plant effluents prior to discharge.

If less stringent disinfection is required, existing chlorination systems may be upgraded by providing higher chlorine dosages, improved mixing and/or longer contact periods. In most cases, improved mixing and longer contact times have proven more effective than increased chlorine dosage. Effective contact time can be increased by improving the flow pattern through the existing basin to eliminate short circuiting or by adding additional tankage.

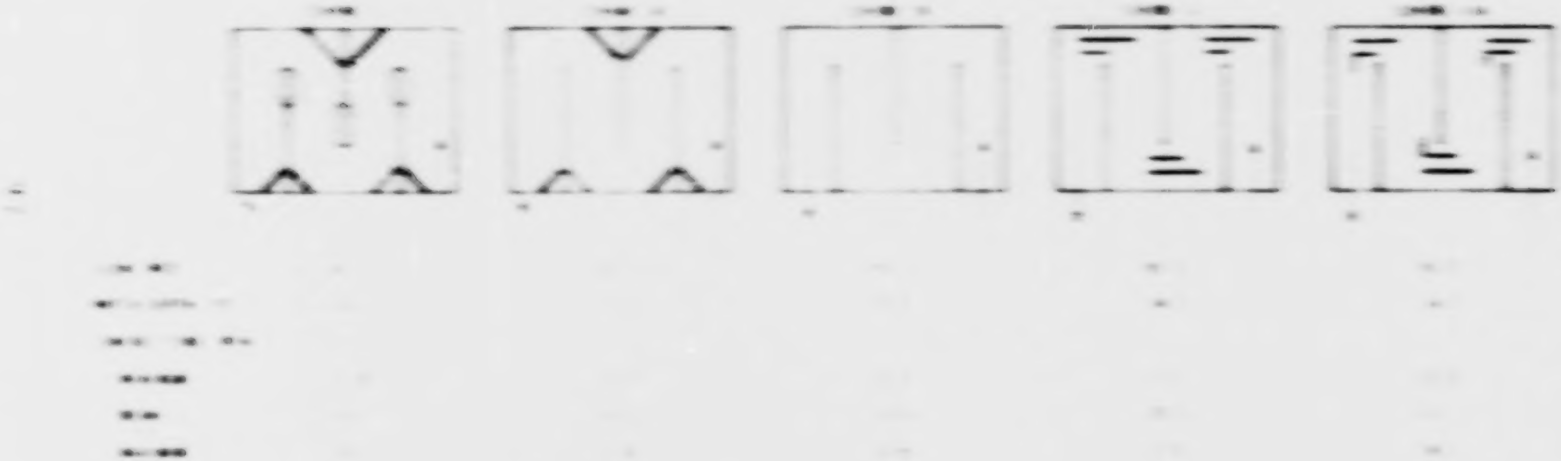
The configuration of the contact tank may result in appreciable differences between the actual and theoretical contact times. Model tests were made by the Metropolitan Sanitary District of Greater Chicago (3) to evaluate the impact of different baffle designs on actual detention time. The various baffle designs evaluated and the test results are shown on Figure 9-3.

The data show that the baffle arrangement used in Scheme IIA, using turning vanes, provided substantially higher actual contact times than the other schemes tested. The improper use of turning vanes, Scheme II, resulted in the lowest actual contact times. Based on the findings of this study, the features of Scheme IIA were incorporated in the design of the 330 mgd chlorination facility at the Chicago Calumet Sewage Treatment Plant. The findings of this study saved the Metropolitan Sanitary District of Greater Chicago an estimated $200,000 in construction costs and considerable reduction in annual costs due to more efficient chlorine mixing.

Rapid dispersement of chlorine at the addition point increases chlorine contact and improves disinfection efficiency. Baffles can be designed to create turbulence at the chlorine addition point and improve mixing. Baffled systems have the advantage of not requiring mechanical equipment. Mechanical mixing or air agitation can be advantageously employed where plant hydraulics will not allow the use of baffles, or where a portion of the existing basin can be converted to a mixing chamber and the remainder of the basin and/or a long outfall sewer can be used to provide the needed contact time.

FIGURE 6.2

IMPACT OF CHLORINE TANK BAFFLE DESIGN ON ACTUAL DETENTION TIME [illegible]



Studies by Kothandaraman and Lin (6) have shown that chlorinated wastewater can be subjected to air agitation with no apparent loss of total chlorine residual. Their studies also indicated a better bacterial kill with air agitation than without air agitation. In those facilities where postaeration facilities are required to raise the DO content of the wastewater, it may be feasible to use the chlorine contact tanks for both postaeration and disinfection.

9.2.4 Chlorination Systems

The most common form of chlorine utilized in the field of wastewater treatment is liquid chlorine. In a typical installation, liquid chlorine is delivered in pressurized containers. Standard sizes for containers are 100-pound, 150-pound and 1-ton cylinders, and 16-ton, 30-ton and 55-ton tank cars. Normally, gaseous chlorine is released from the top of the container and is drawn through the chlorinators by a vacuum. In larger plants where the evaporation rate within the container cannot meet the demand, liquid chlorine is withdrawn and evaporated in separate evaporators. The chlorine gas is then mixed with plant effluent or potable water (small plants) to form a chlorine solution which is fed into the wastewater. Such systems have proven to be highly reliable and economical. However, there are ramifications that should be considered:

1. Liquid chlorine is a potentially dangerous material.

2. The demand for liquid chlorine has been increasing steadily, while the construction of liquid chlorine manufacturing facilities has lagged.

3. Unless the above trend is corrected, it can be expected that shortages of liquid chlorine will develop, and the cost will significantly increase.

In recent years, the utilization of sodium hypochlorite for disinfection has become attractive. Sodium hypochlorite can either be delivered to the site in liquid form in 500 to 5,000 gallon tank cars or trucks, or manufactured on site. When delivered to the site, it is normally sold at a concentration of 12 to 15 percent by weight of available chlorine. The tank cars are unloaded into storage tanks, and the hypochlorite is fed to the wastewater through chemical feeders. Sodium hypochlorite can be manufactured on site from salt or from seawater. Generation units are now being produced by at least four major manufacturers. While most of these units produce hypochlorite at concentrations of one percent or less, one system produces hypochlorite at a concentration of about eight percent.

9.2.5 Ozonation

Based on the success of ozonation in the water treatment field, ozone has been considered as a possible disinfectant for wastewater. Some of the advantages of ozonation are as follows:

1. Ozone is generated from air and its supply is dependent only on a source of power. Ozone can also be produced from oxygen at significant energy savings, an option which should be considered where on-site generation of oxygen is practiced in conjunction with biological treatment.

2. Ozone may be a more efficient viricide than chlorine.

3. Ozonation can also be used as a tertiary treatment process for oxidation of residual carbon compounds and for odor control.

4. The Maximum Allowable Concentration (MAC) of ozone in air as established by the American Council of Governmental Industrial Hygienists is 0.1 ppm by volume for continuous human exposure. The threshold odor of ozone is 0.01 to 0.02 ppm. This means that a person working near an ozone handling area should be able to detect the presence of ozone at concentrations far below the MAC.

5. Ozonated effluents have not been shown to be toxic to the receiving water biota, as have residual chlorine compounds such as chloramines and chlorinated hydrocarbons. Ozone does not increase the dissolved solids concentration in the effluent as does chlorine.

6. Ozonation of effluents increases the DO.

The predominant disadvantage of ozonation is the high capital and operating cost associated with its generation. Further development of the process may lead to better economy which would encourage greater use of ozonation for wastewater disinfection.

Nebel, et al. (7) conducted pilot plant studies at Louisville, Kentucky, using ozone for disinfection of activated sludge effluent. The wastewater contained a significant amount of nonbiodegradable industrial organic wastes which exerted an above-normal ozone demand. Nevertheless, the tests indicated that a dosage of 15 mg/l and a contact time of 22 minutes resulted in excellent destruction of fecal and total coliforms and fecal streptococci.

During 1973, two separate cost evaluations of alternative disinfection systems were made by Metcalf & Eddy, Engineers in Boston. Both of the plants studied were large; one a 105 mgd plant located on the eastern seaboard, and the other a 240-mgd plant located in the midwest. The eastern plant study included cost comparisons of liquid chlorine, trucked-in and on-site generated sodium hypochlorite and ozonation. The midwestern study compared all of the above alternatives except ozonation. The results of the two studies are presented in Table 9-3. The costs are shown as a ratio to the liquid chlorine cost and are a sum of the annual operating costs and amortized capital costs.

TABLE 9-3
RELATIVE TOTAL ANNUAL COSTS OF DISINFECTION ALTERNATIVES

Alternative	Cost Ratio[1] Eastern Plant 105 mgd	 Midwestern Plant 240 mgd
Liquid Chlorine	1	1
Sodium Hypochlorite, Trucked to Site	2.0	2.4
On-Site Generation		
0.8% Sodium Hypochlorite	1.2	1.4
15% Sodium Hypochlorite	1.8	2.1
Ozone (Generated from Oxygen)	4.0	

[1] Data used to compute cost ratios:

1. Cost of liquid chlorine	$88.19/ton	$75/ton
2. Cost of liquid sodium hypochlorite	$0.18/gal	$0.30/gal
3. Capital cost of on-site generator (15%)	$4,300,000	$1,116,000
4. Capital cost of on-site generator (0.8%)	$2,900,000	$ 975,000
5. Average ozone dosage	10 mg/l	
6. Average chlorine dosage	16.6 mg/l	5 mg/l
7. Amortization rate	10 yr. at 6½%	15 yr. at 7%

9.4 Odor Control

Wastewater treatment plants serving large municipalities are generally characterized by extensive collection systems with correspondingly high detention times. For example, the Washington, D.C. Pollution Control Plant serves areas as far as 25 to 30 miles away. This type of situation often leads to odor problems during summer periods. Odor problems are characteristically most critical during the plant's low flow periods (approximately 9 PM to 4 AM), due to increases in the sewer detention time.

9.4.1 Odor Generation and Characteristics

Odors from wastewater treatment plants can usually be attributed to three sources: septic raw wastewater, overloaded secondary treatment facilities and sludge treatment practices.

Septicity in wastewaters is caused by the depletion of DO due to long residence in sewers and the subsequent increase in anaerobic activity. As wastewater becomes septic, facultative and anaerobic bacteria flourish. These bacteria utilize nitrates and sulfates present in wastewater as their oxygen source. The reduction of sulfate ions produces the highly odorous gas, hydrogen sulfide. Other odorous gases which may be present are indole, skatole, mercaptans, disulfides, volatile fatty acids and ammonia.

Increased summer temperature and extended sewer detention times can result in the rapid buildup of hydrogen sulfide and carbon dioxide according to the following reactions (8):

$$SO_4^{=} + \text{organics} \xrightarrow[\text{bacteria}]{\text{anaerobic}} S^{=} + H_2O + CO_2$$

$$S^{=} + 2H^{+} \longrightarrow H_2S$$

At a pH level below 8, the equilibrium shifts toward the formation of nonionized H_2S and is about 80 percent complete at pH 7. At pH 8 and above, most of the reduced sulfur exists in solution as HS^- and $S^=$ ions (8). H_2S is noticeable even in the cold when present in water to the extent of 0.5 mg/l. When present to the extent of 1.0 mg/l, it becomes very offensive (9).

Overloaded secondary treatment facilities are also a potential source of odor. If the air supply to an activated sludge aeration tank is inadequate, odorous conditions usually develop. It is also possible that a properly sized air supply system can strip odorous gases from septic wastewater.

Odors associated with sludge treatment occur in thickening, digestion and sludge dewatering facilities. Thickeners may receive both septic primary and secondary sludges. Gases from well-operated digesters may contain small quantities of H_2S, which are usually destroyed by normal flaring of digester gas. The predominant odor in digested sludge is ammonia, although traces of volatile organic acids may be present.

9.1.2 Odor Measurement

Odor data are generally qualitative rather than quantitative in nature. The two available quantitative methods are the H_2S determination and the Threshold Odor Number (TON). The latter method is only semiquantitative in that determination of the TON is dependent on the olfactory senses of the individual performing the analysis. This can be depersonalized somewhat by using a panel to determine the TON value for a sample.

9.3.3 Odor Control Methods Available

The various methods available for control of odor emanating from a wastewater treatment plant are:

1. Changes in the operational procedures and new techniques.
2. Chemical treatment or pretreatment, which might include chlorine, ozone, lime or powdered carbon.
3. Collection and treatment of noxious gases.

9.3.3.1 Changes in Operational Procedures and New Techniques

Odors associated with septic wastewater are generally not amenable to solution through operational changes within the treatment plant itself. These odors must be controlled upstream of the plant through the use of aeration or chemical treatment methods. The applicability of in-sewer aeration methods for reduction of odors and hydrogen sulfide corrosion is discussed in the Process Design Manual for Sulfide Control in Sanitary Sewerage Systems (10). Among the procedures that have been evaluated are U-tube installations (refer to Section 8.2.1.4) and pure oxygen injection into force mains.

Many sludge odors in a plant are a direct result of an improperly operated or overloaded anaerobic sludge digester. Improved temperature control and better mixing of digester contents may alleviate the odor problem.

9.3.3.2 Chemical Treatment of Wastewater

Chlorination is probably the most widely used of the chemical treatment processes available to control odors for the following reasons:

1. It is highly effective.
2. Many treatment operators have had experience in handling chlorine.
3. Chlorination facilities already exist at most plants for disinfection.

Chlorination is used for two purposes: to retard biological action, which produces odors, and to react chemically with odorous sulfur compounds, oxidizing them to innocuous sulfur forms, usually free colloidal sulfur.

Table 9-3 contains a summary of odor reduction data for a chlorinated raw domestic wastewater (11). A prechlorination dosage of 10 mg/l at maximum flow was recommended for odor control. In addition, another incremental 5 mg/l of chlorine capacity was recommended as an adequate safety margin for peaks in sulfide levels or chlorine demand.

TABLE 9-3
EFFECT OF CHLORINE ON ODOR REDUCTION FOR A RAW DOMESTIC WASTEWATER (11)

Chlorine[1] Dosage mg/l	Detention Time in Minutes	
	1	130
	Sulfide mg/l	Sulfide mg/l
0	1.7	1.7
5	0.5	0.7
10	0.2	0.3
25	0.2	0.2
50	0.2	0.2

[1] pH = 7.
Air Temperature = 65 deg F.

Ozone has been added to wastewaters for odor control with some favorable results. Because of the extremely high reactivity of ozone, a much higher ozone demand is generally exhibited by a wastewater than would be exhibited for chlorine. However, the use of pure oxygen in activated sludge treatment may have an added benefit, since the oxygen gas could provide the ozone generator with an economic source of oxygen. Due to the high cost of ozone generation, the use of ozone for odor control may be limited (11).

Lime and powdered carbon have also been used in various applications for odor control. The addition of lime to septic wastewater raises the pH. Since the solubility of H_2S increases with increasing pH, less H_2S evolves, thereby decreasing the odor level. Powdered activated carbon adsorbs odor causing materials and thereby decreases the odor level. The results of a laboratory odor study are presented in Table 9-4 (12). Concentrations of less than 10 mg/l of powdered activated carbon were successful in providing significant odor reduction. Additional design information for controlling odors is available in the Process Design Manual for Sulfide Control (10).

TABLE 9-4
EFFECT OF POWDERED ACTIVATED CARBON
ON ODOR REDUCTION (12)

Plant	Raw Domestic Wastewater[1] TON[2]	Activated Carbon Treated Effluent TON[2]	Concentration of Activated Carbon mg/l
Charlottesville, Va.	400	100	4.0
Hershey Estates, Pa.	140	120	5.0
Butler, Pa.	300	300	10.0

[1] Sample temperature = 60 deg F.
[2] Threshold Odor Number.

9.3.3.3 Collection and Treatment of Noxious Gases

The covering of odorous unit process facilities to localize odors is a method which can be used to prevent odors from reaching the atmosphere. The major expenses of this method are the covering of the units and collection and treatment of the evacuated gases. In cold climates, covering units can lead to conditions of high humidity and indoor fog if proper ventilation is not provided. Many municipal plants, e.g., Cedar Rapids, Iowa (13), and Elmira, New York (14), are using low cost, formed-in-place styrofoam domes on odorous treatment units.

The treatment methods usually considered for evacuated gases include simple or catalytic combustion, ozonation and chemical oxidation. Combustion methods require heating the gases to approximately 800 to 900 deg F for catalytic combustion and approximately 1,300 to 1,400 deg F for simple combustion. Operating costs for these methods are primarily determined by the amount of air to be heated. Ozonation costs, while somewhat affected by the volume of gas collected, are primarily affected by the quantity of odorous materials to be controlled.

9.3.4 Effect on Subsequent Units

A consideration in using chlorine for odor control is that the chlorine dosage should not produce a high residual chlorine level which may in turn be detrimental to secondary biological units. When using lime for odor control, consideration must also be given to increased sludge production.

9.3.5 Process Designs and Cost Estimates

A cost estimate has been prepared for two odor control systems (chlorination and powdered activated carbon) for 1, 3 and 5 mgd treatment plants. The capital costs are presented in Table 9.5. The chemicals are added to raw wastewater before the downstream treatment units.

TABLE 9.5
CAPITAL COSTS FOR ODOR CONTROL SYSTEMS

Plant Size mgd	Capital Costs for Odor Control Systems	
	Chlorination 10 mg/l	Powdered Activated Carbon 10 mg/l
1	\$46,000[1]	\$37,000
3	49,000[1]	52,000
5	51,000[1]	73,000

[1] Smallest size commercially available chlorinator.

The chlorination system included a gas chlorinator capable of delivering 10 mg/l of chlorine during peak flow rates. A building, scale and other necessary appurtenances were included. The powdered activated carbon system included a 15 day storage hopper and a volumetric feeder capable of delivering 10 mg/l at peak flow rates. In addition, a one-day capacity slurry tank, pump, building and associated piping were included.

9.4 Other Uses of Chlorine

In the operation of wastewater treatment plants, chlorine has been found useful as an upgrading technique. Some of the various applications of chlorine are as follows (4) (14) (15):

1. Destruction or control of undesirable growths of algae and slime forming bacteria in pipelines and conduits.

2. Control of filter flies, clogging and ponding in trickling filters (Chlorine applied for approximately 8 hours to produce a residual of 1 to 2 mg/l in the distributor arm will generally unclog the filter. Residuals of 20 to 50 mg/l will eliminate ponding by causing the filter to unload all of its biological slime).

3. Improvement in wastewater coagulation

4. Improvement in the separation of grease from wastewater

5. Reduction of the immediate oxygen requirements of return activated sludge and digester supernatant return

6. Chlorination of return activated sludge to control sludge bulking (A chlorine dose of 1 to 10 mg/l based on return sludge flow has been used, provided the chlorine application point is located to allow for 2 to 3 minutes of mixing before being discharged to the aeration basin; otherwise high "spot chlorine" concentrations may damage the activated sludge biomass)

9.5 References

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12. *Aqua Nuchar for Odor Control in Waste Treatment.* Covington, Virginia.

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CHAPTER 10

SLUDGE THICKENING

10.1 Sludge Treatment

Raw sludge is normally unsuitable for disposal without prior treatment. Such treatment is required to reduce the volume to be handled and to provide stabilization. Sludge treatment is the most complex and costly aspect of wastewater treatment. Although the volume of sludge produced is frequently less than one percent of the total volume of wastewater treated, sludge treatment facilities can often account for 25 to 50 percent of the total capital and operating costs for the entire plant (1).

10.1.1 Sludge Treatment Processes

The processes used for sludge treatment may be grouped into the following categories:

1. Thickening
2. Stabilization
3. Dewatering

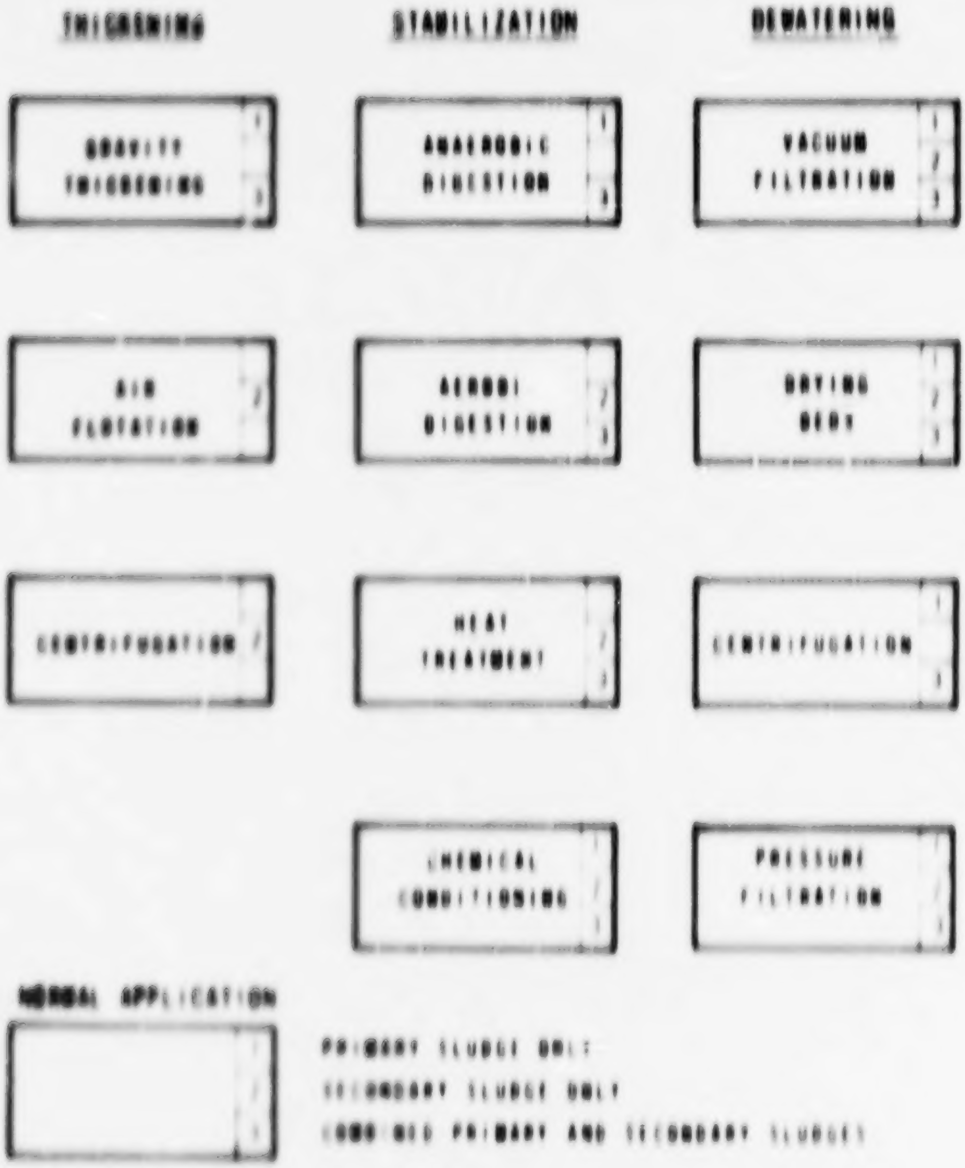
The important processes included in each of these categories are shown on Figure 10.1.

The separation of solids in either primary or secondary sedimentation tanks has two objectives. These are the recovery of a high percentage of solids from the liquid portion as required by effluent standards, and the concentration of these solids to reduce the required capacity of the sludge processing units (2). These objectives are conflicting in that highest effluent quality is achieved by immediate removal of settled solids, without allowing them to accumulate in sedimentation tanks. Although primary sludges can often be effectively thickened in the sedimentation tanks, it is preferable to remove secondary sludges quickly and provide separate thickening units. Thickening of sludge to reduce its volume prior to stabilization, dewatering, and ultimate disposal in fact generally provides the highest benefit to cost ratio of all sludge handling processes(2). This chapter is directed to various methods of sludge thickening.

Sludge stabilization processes are used to convert raw primary and secondary sludges to inoffensive forms by reducing the organic material in the sludges or by otherwise rendering them inert. This is especially essential when sludge is to be disposed of on land. Sludge dewatering characteristics may also be improved by some methods of stabilization such as thermal or lime conditioning. Where incineration is to be used for ultimate disposal, organic

FIGURE 10-1

SLUDGE TREATMENT PROCESSES


THICKENING
STABILIZATION
DEWATERING
GRAVITY THICKENING
1
3
ANAEROBIC DIGESTION
1
3
VACUUM FILTRATION
1
2
3
AIR FLOTATION
2
AEROBIC DIGESTION
2
3
DRYING BEDS
1
2
3
CENTRIFUGATION
2
HEAT TREATMENT
2
3
CENTRIFUGATION
1
3
CHEMICAL CONDITIONING
1
2
3
PRESSURE FILTRATION
1
2
3
NORMAL APPLICATION
1
2
3
PRIMARY SLUDGE ONLY
SECONDARY SLUDGE ONLY
COMBINED PRIMARY AND SECONDARY SLUDGES

matter is required for self-supporting combustion and no prior organic reduction process should be used. Sludge stabilization is considered in detail in Chapter 11.

Sludge dewatering is advantageous because it reduces the volume of sludge that will require further processing and ultimate disposal. Where the sludge must be transported to a disposal site, dewatering reduces the land requirements as well as handling costs. Dewatering prior to incineration is necessary to remove excess water that otherwise would have to be removed in the incinerator at the expense of auxiliary fuel. Sludge dewatering is the subject of discussion in Chapter 12.

10.1.2 Sludge Processing Interrelationships

The various unit processes shown on Figure 10-1 may be combined in numerous different sludge treatment configurations. Typical flow diagrams for several commonly used sludge treatment systems are shown on Figures 10-2 and 10-3. The systems have been grouped into two general categories, depending on whether or not biological sludge stabilization is involved. Systems using biological stabilization are shown on Figure 10-2 and nonbiological systems are shown on Figure 10-3.

The selection of the individual components of a sludge treatment system must not be made independently of the other components to be used. Instead, a cost-effective system of compatible sludge thickening, stabilization, and dewatering units must be selected on the basis of the local situation. The method to be used for ultimate disposal of the sludge after treatment may limit the selection of a sludge treatment system for a particular application. The alternative process units that can be considered will depend on the type of wastewater treatment, the sludge characteristics, and the space available for the treatment facilities.

Small plants are generally located in less densely populated areas where land availability is not a primary constraint for either treatment facilities or ultimate sludge disposal and where skilled operators may be difficult to obtain. In these instances a simple system such as is shown on Figure 10-2(a) may be adequate. Stabilized sludge in this case is dried on sand beds prior to ultimate disposal. Plants serving populations larger than 10,000 often find mechanical devices more economical for sludge dewatering. A system using mechanical sludge dewatering devices is shown on Figure 10-2(b). In this situation, waste activated sludge is returned to the plant influent and the combined sludge is concentrated in the primary clarifier without separate thickeners. Where separate thickening of waste activated sludge is desirable prior to stabilization, a system such as shown on Figure 10-2(c) is commonly used.

Nonbiological sludge processing systems are used mainly when:

1. The nature of the sludge makes it difficult to digest

FIGURE 10.2

TYPICAL SLUDGE PROCESSING SYSTEMS USING BIOLOGICAL STABILIZATION (3)

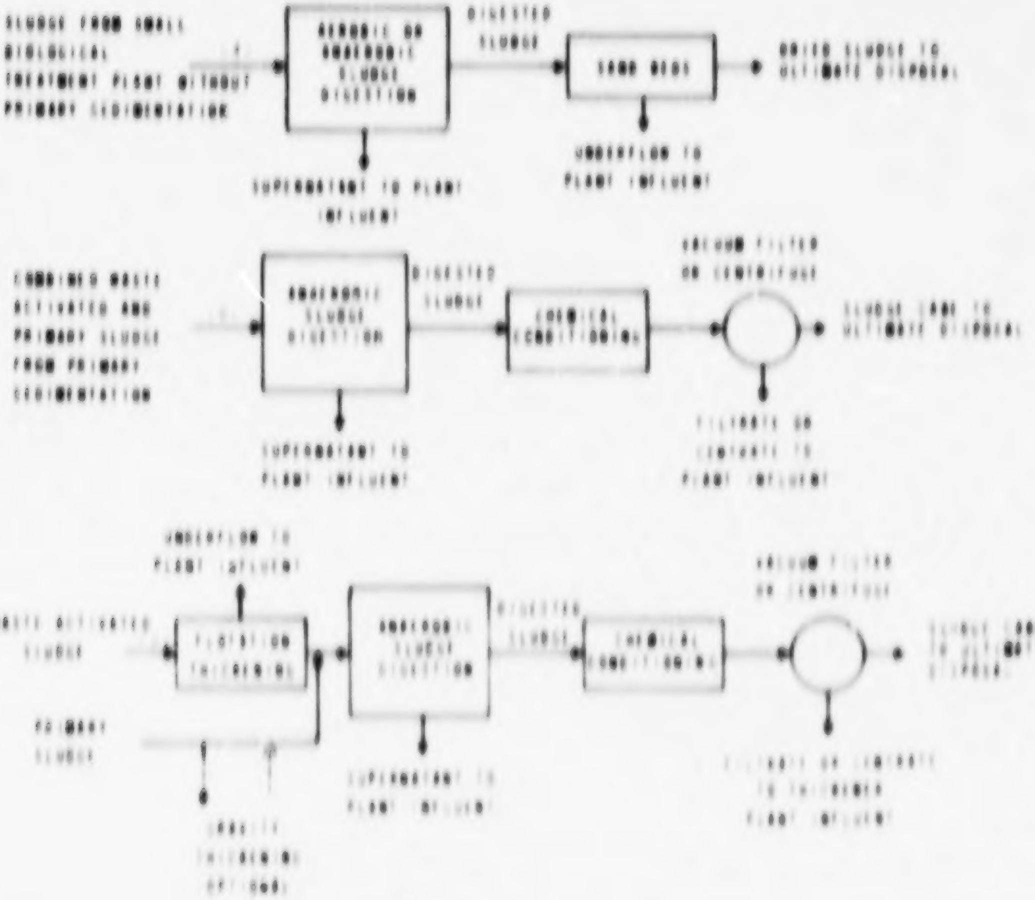
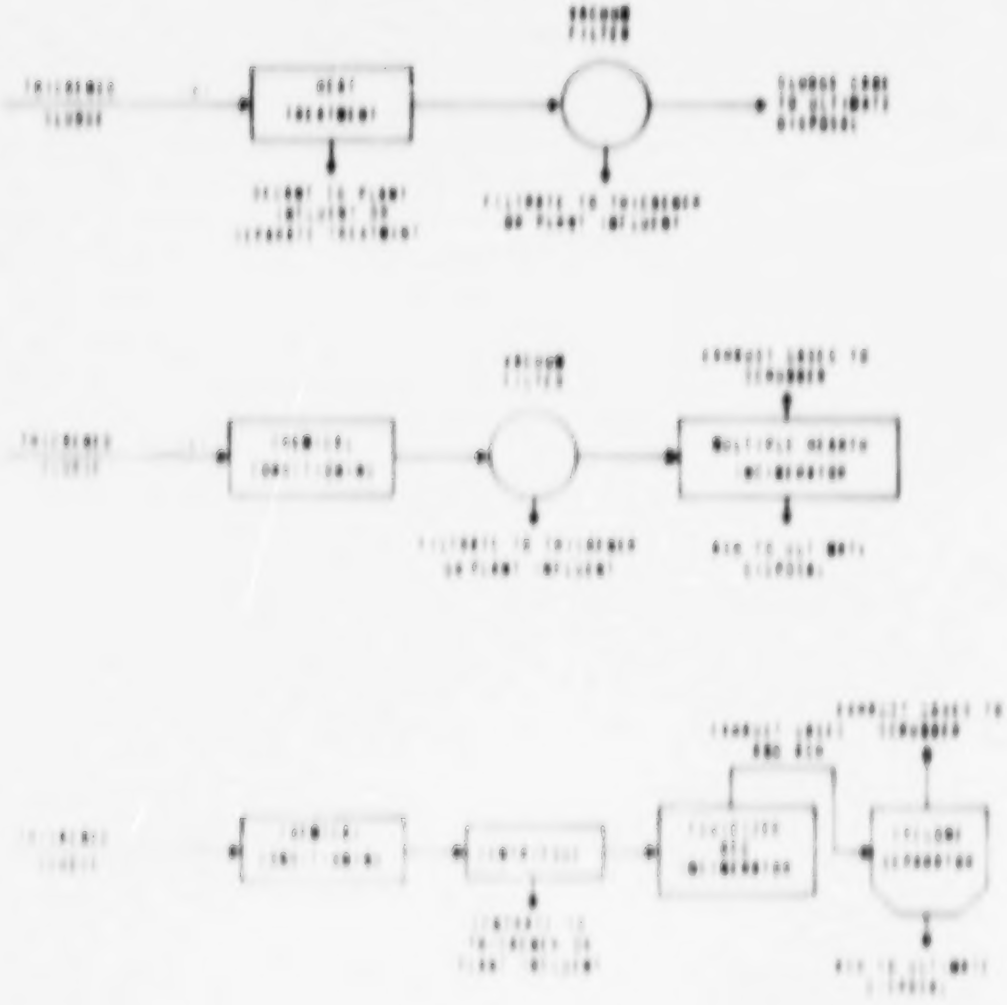

SLUDGE FROM SMALL BIOLOGICAL TREATMENT PLANT WITHOUT PRIMARY SEDIMENTATION
AEROBIC OR ANAEROBIC SLUDGE DIGESTION
DIGESTED SLUDGE
SAND BEDS
DRIED SLUDGE TO ULTIMATE DISPOSAL
UNDERFLOW TO PLANT INFLUENT
SUPERNATANT TO PLANT INFLUENT
COMBINED WASTE ACTIVATED AND PRIMARY SLUDGE FROM PRIMARY SEDIMENTATION
ANAEROBIC SLUDGE DIGESTION
DIGESTED SLUDGE
CHEMICAL CONDITIONING
VACUUM FILTER OR CENTRIFUGE
SLUDGE CAKE TO ULTIMATE DISPOSAL
SUPERNATANT TO PLANT INFLUENT
FILTRATE OR CENTRATE TO PLANT INFLUENT
UNDERFLOW TO PLANT INFLUENT
WASTE ACTIVATED SLUDGE
FLOTATION THICKENING
ANAEROBIC SLUDGE DIGESTION
DIGESTED SLUDGE
CHEMICAL CONDITIONING
VACUUM FILTER OR CENTRIFUGE
SLUDGE CAKE TO ULTIMATE DISPOSAL
PRIMARY SLUDGE
GRAVITY THICKENING (OPTIONAL)
SUPERNATANT TO PLANT INFLUENT
FILTRATE OR CENTRATE TO THICKENER PLANT INFLUENT

FIGURE 10-3

TYPICAL SLUDGE PROCESSING SYSTEMS USING NONBIOLOGICAL STABILIZATION (3)


VACUUM FILTER
THICKENED SLUDGE
HEAT TREATMENT
SLUDGE CAKE TO ULTIMATE DISPOSAL
DECANT TO PLANT INFLUENT OR SEPARATE TREATMENT
FILTRATE TO THICKENER OR PLANT INFLUENT
VACUUM FILTER
EXHAUST GASES TO SCRUBBER
THICKENED SLUDGE
CHEMICAL CONDITIONING
MULTIPLE HEARTH INCINERATOR
FILTRATE TO THICKENER OR PLANT INFLUENT
ASH TO ULTIMATE DISPOSAL
EXHAUST GASES TO SCRUBBER
EXHAUST GASES AND ASH
THICKENED SLUDGE
CHEMICAL CONDITIONING
CENTRIFUGE
FLUIDIZED BED INCINERATOR
CYCLONE SEPARATOR
CENTRATE TO THICKENER OR PLANT INFLUENT
ASH TO ULTIMATE DISPOSAL

2. A chemically or thermally conditioned sludge, dewatered by filtration or centrifugation, is acceptable for land disposal and is the most economical alternative.

3. Incineration is necessary for maximum sludge volume reduction thereby making it uneconomical to provide anaerobic digestion, due to the loss of caloric value.

Nonbiological systems such as shown on Figure 10-3 may be used in these instances. Heat treatment may be used to stabilize sludge organics prior to dewatering and ultimate disposal as shown on Figure 10-3(a). Incineration systems such as shown on Figures 10-3(b) and 10-3(c) provide the highest degree of sludge volume and organic reduction. Although the capital and operating costs of these units are high, the site requirements for disposing of the residue are considerably reduced.

10.2 General Sludge Thickening Considerations

Sludge thickening is commonly used as the first step in a sludge treatment system to separate more water from the sludge solids than can be accomplished in the wastewater clarifiers. This reduces the required capacity and increases the efficiency of subsequent sludge stabilization and dewatering processes that have higher unit costs than thickening. The processes commonly used to thicken sludge are gravity thickening, air flotation thickening, and centrifugation.

Gravity thickeners are used extensively since they are the most economical method of thickening primary sludge or some blended primary and waste activated sludges. Gravity thickeners are particularly applicable to smaller installations where the thickened sludge will be anaerobically digested and dried on sand beds. Although waste activated sludge is difficult to thicken, it often is effectively handled by mixing with primary solids prior to gravity thickening. Where the weight of waste activated sludge is greater than 40 percent of the total sludge weight, gravity thickening is not effective and separate thickening of waste activated sludge by other methods should be considered.

Air flotation is a commonly used method of thickening for large quantities of waste activated sludge. Under these circumstances, primary sludge can be thickened in the primary clarifiers or by separate gravity thickening. Normally, air flotation thickening in conjunction with separate primary sludge thickening will produce the highest overall sludge concentration.

Centrifugation is a very adaptable process capable of thickening sludges with a wide range of characteristics. Due to high operating, maintenance and power requirements, centrifugal thickening is generally used only where special limitations of sludge characteristics make other methods unsuitable.

An important consideration in the use of any sludge processing unit is the effect of recycled flows that are returned to the head of the plant. The quality of the overflow from well operated gravity thickeners and the underflow from flotation thickeners is similar to that of raw wastewater (1). However, return flows from thickeners are normally of sufficiently large volume to result in recycled BOD and SS loads which can adversely affect overall plant performance. With centrifugal thickening, high solids captures must be obtained to avoid recycle of unsettleable fine solids to the head of the plant, which will adversely affect plant effluent quality.

Thickening is an operation that produces a more concentrated and therefore more viscous sludge. The more viscous sludge will result in higher friction losses than water or dilute sludge, and it is important that the thickened sludge pumping and piping system be designed to handle the sludge throughout the range of expected solids concentrations. Because the consistency of the sludge may vary under different conditions, pumps must be able to operate properly over a wide range of discharge pressures. The design of sludge piping systems is complicated by the need for both large sizes to decrease the likelihood of stoppages and to facilitate cleaning, and for high velocities to minimize the deposition of grease and the settling of solids in the pipes. To minimize friction losses, sludge lines should be as short as possible and should contain a minimum of bends and fittings.

10.3 Gravity Thickening

Gravity thickening is the most common process in use today for the concentration of sludge prior to digestion and/or dewatering. Thickeners can contribute to the upgrading of sludge handling facilities as follows:

1. Increase the hydraulic capacity of overloaded digesters or subsequent sludge handling units.

2. Improve primary clarifier performance by providing continuous withdrawal of sludge, thereby ensuring maximum removal of solids.

The process is simple and is the least expensive of the available thickening processes. The reduction in size and improvement in efficiency of subsequent sludge handling processes often can offset the cost of gravity thickening. The process also allows equalization and blending of sludges, thereby improving the uniformity of feed solids to the following processes. Existing gravity thickeners can be upgraded by providing continuous feed and drawoff, by diluting the feed solids, and by chemical addition.

10.3.1 Process Considerations

Gravity thickening is characterized by zone settling. The four basic settling zones in a thickener are:

1. The clarification zone at the top containing the relatively clear supernatant.

2. The hindered settling zone where the suspension moves downward at a constant rate and a layer of settled solids begins building from the bottom of the zone.

3. The transition zone characterized by a decreasing solids settling rate.

4. The compression zone where consolidation of sludge results solely from liquid being forced upward around the solids.

To date, many attempts have been made to simulate zone settling in a batch settling test to generate design information which would be applicable to a continuous unit. Various theories have been developed for analyzing batch settling data and they have been reviewed and discussed in the literature (4) (5) (6) (7). Most of the theories assume that the settling velocity of sludge at a given concentration in a small batch cylinder is similar to the velocity in prototype thickening units. However, it has also been recognized that other parameters are involved, such as cylinder depth, cylinder diameter, mixing conditions, and sludge characteristics. All have a definite influence on thickening performance. The cumulative effect of these parameters is such that when batch settling test data are used for unit sizing, the result is an oversized unit. For this reason, batch settling test results must be used with caution (5). Edde et al (6) have developed a mathematical model from batch and full scale thickening data. The model facilitates determination of solids loading at a given sludge blanket depth, initial feed solids, underflow concentration, and hydraulic loading. This technique is particularly useful for determining gravity thickener design parameters when upgrading existing wastewater treatment plants.

10.3.2 Design Considerations

Both solids and hydraulic surface loadings must be considered when designing gravity thickeners. Experience indicates that solids loading generally governs the design (8). Design solids loadings and expected underflow concentrations are shown in Table 10-1.

The dry solids ratio of waste activated to primary sludge governs the acceptable solids loading to be used in thickener design. As this ratio increases, the acceptable solids loading decreases.

Most thickeners are operated at a hydraulic loading of 600 to 800 gpd/sq ft (9). Thickeners with hydraulic loadings less than 400 gpd/sq ft have been found to produce odors (9). To achieve hydraulic loadings in the acceptable range, secondary effluent is normally blended with the combined waste sludge before feeding the resulting uniform diluted sludge to the thickener.

TABLE 10-1
THICKENER DESIGN LOADINGS AND UNDERFLOW CONCENTRATIONS

Type of Sludge	Underflow Concentration % TS	Average Solids Loading psf/day
Primary	8-10	20
Primary and Trickling Filter	6-8	12
Primary and Waste Activated (60:40 Weight Ratio)	3-6	8

As mentioned previously, solids loading is generally the controlling parameter and dictates the required surface area of the thickener. For example, if a solids loading of 10 psf/day is used for a combined primary and activated sludge and typical performance efficiencies are desired, the calculated hydraulic loading will be in the order of 25 to 50 gpd/sq ft. Dilution will be required to achieve the recommended 600 to 800 gpd/sq ft.

An important consideration in designing a thickener is to provide the capability to handle peak conditions. Gravity thickeners are normally sized for average sludge production rates. Peak sludge production rates often are adequately handled by storage available within the system, either in the primary tanks or the thickener itself. However, a solids balance for the system as it will be operated should be determined to indicate whether supplementary storage or additional thickener capacity is required to accommodate the peak conditions. Sufficient storage normally should be available within the system for the maximum three day plant solids loading.

Most continuous thickeners today are circular and designed with a side water depth of approximately 10 feet. While sludge blanket depth is an important parameter, it has been reported that underflow solids concentrations are independent of sludge blanket depths greater than 3 feet (1). Increased sludge detention time in the thickener will result in increased underflow solids concentration. A detention period of 24 hours has been suggested as the time required to achieve maximum compaction (1). Sludge blanket depth and detention time are closely interrelated. The sludge blanket depth may be varied with fluctuation in solids production to achieve good compaction. During peak conditions the detention time may have to be shortened to keep the sludge blanket depth sufficiently below the overflow weirs to prevent excessive solids carry over.

The performance of a gravity thickener depends a great deal upon the type of sludge to be thickened. Generally, gravity thickening should not be used for thickening activated sludge alone.

In the design of gravity thickeners, it is important that operational flexibility be provided. Such flexibility includes the ability to regulate the quantity of dilution water, adequate sludge pumping capacity so that solids concentrations are not limited, continuous feed and underflow pumping, protection devices against torque overload, and a sludge blanket detection device.

10.3.3 Use of Gravity Thickening for Upgrading Existing Sludge Handling Facilities

Gravity thickening is often used prior to digestion processes. It can also be used as a combined thickening and equalization process prior to sludge dewatering. Another application is in areas where sludge hauling is utilized and there is a need to reduce the volume of sludge to be hauled. In all cases, gravity thickening will yield higher underflow solids concentrations than obtainable with primary sedimentation, and the efficiency of subsequent digestion and dewatering facilities will improve. Hence, gravity thickening should always be considered in upgrading existing solids handling and dewatering facilities.

10.3.4 Upgrading Existing Gravity Thickeners

Improved thickening can be obtained by diluting the sludges to be thickened. It has been reported that a feed solids concentration of 0.5 to 1.0 percent is optimum and that dilution reduces the interference between the settling particles (1).

Torpey (10) used dilution for thickening combined primary and secondary sludges in the development of the Densludge System. A feed sludge concentration of less than 1 percent produced underflow concentrations of 11.2 and 6 percent for combined primary and modified waste activated sludge, and combined primary and conventional waste activated sludge, respectively. To obtain these dilute feed sludge concentrations, dilution water was pumped from the primary or secondary clarifier and blended with the combined sludges prior to thickening.

Thickening systems at New York City's Tallmans Island and Bowery Bay pollution control facilities utilize the processes developed by Torpey and presently obtain 4 to 6 percent underflow solids concentration with a yearly average of 4.5 percent (11). Both plants are operating using a combination of the step and activated aeration processes. A similar design with digested sludge recirculation at Bergen County, New Jersey, produces an underflow concentration of 5.2 to 7.5 percent with a yearly average of 6.3 percent (12). In both cases, the lower underflow concentrations occur during the summer months.

The improved thickening due to dilution can also be attributed to the fines that are washed from the sludge and returned to the plant through the thickener overflow. Experiences at Bergen County, New Jersey, show that, even with digested sludge recirculated to the thickener, the overflow from the thickener does not appreciably affect the overall BOD removal efficiency of the treatment plant (13).

Similar experiences were reported at the Bowery Bay Pollution Control Plant. However, the air requirements were increased from 0.28 to 0.54 cu ft/gal (14) by the return of thickener supernatant to the aeration tanks.

Experiences with the use of polyelectrolytes for upgrading gravity thickening have been reported at Amarillo, Texas (15). The original plant was designed to handle 7.5 mgd, but was receiving a flow of up to 10.5 mgd. A 55 foot diameter gravity thickener was being used to thicken combined primary and waste activated sludge. However, bulking occurred in the thickener due to the overloaded conditions and the low ratio of primary sludge to waste activated sludge. In-plant recycling of solids resulted. To minimize the problem, only activated sludge was thickened in the gravity thickener. Primary sludge was thickened in the primary clarifiers. Polyelectrolyte addition was utilized in the thickener in an attempt to improve sludge blanket control and to obtain maximum underflow solids concentration. It was found that a cationic polyelectrolyte permitted a solids loading of 4.5 to 7.0 psf/day while maintaining an underflow solids concentration of 2.6 percent, at a cost of $1.10 to $3.64 per ton of dry solids. Polyelectrolyte was used here to successfully control the sludge blanket height. This practice was continued until the plant was upgraded to 12 mgd.

For the expansion to 12 mgd at Amarillo, Texas, an existing 70 foot diameter final clarifier was modified for thickening the waste activated sludge. The overflow from this thickener is mixed with the primary sludge to dilute the feed sludge to the existing primary sludge thickener (15). Operation of the waste activated sludge thickener showed that at a solids loading of 2 to 4.5 psf/day, only a 2.4 percent underflow solids concentration could be obtained. Polymer addition was tried once again to increase the underflow solids concentration during a 112 day program, but proved unsuccessful.

In Chicago, the addition of polyelectrolyte at dosages of less than 10 lb/ton dry solids increased the solids loading by two to four times, but there was no improvement in underflow solids concentration (16). From these two examples, it appears that polyelectrolyte addition improves solids capture and reduces solids overflow, but has little or no effect on improving solids underflow concentration.

10.1.5 Process Design and Cost Estimates

The following example illustrates the use of gravity thickening before anaerobic digestion.

Existing anaerobic digesters were experiencing unstable operation due to the increased amount of sludge generated by an increase in plant flow from 10 to 16 mgd. Operational data from the overloaded plant without gravity thickening and the upgraded plant with thickening are presented in Table 10-2. The volume of the combined primary and secondary sludge was 100,000 gpd at 3 percent solids prior to upgrading. The increased sludge volume due to the plant overloading decreased the detention time in the digesters from 17 to 11.25 days.

TABLE 10-3
EXAMPLE OF
UPGRADING SLUDGE HANDLING FACILITIES
USING GRAVITY THICKENING

Description	Overloaded Plant	Upgraded Plant
Loadings to the Anaerobic Digester		
Total Weight Primary and Secondary Solids	25,100 lb	25,100 lb
Total Volume Primary and Secondary Sludge	100,000 gpd (3%)	66,200 gpd (4.5%)
Digester Hydraulic Detention Time	11.25 days	17 days

To improve the operation of the existing digesters, it was decided to reduce the sludge volume by gravity thickening. The flow diagram of the upgraded plant is shown on Figure 10-4. The total volume of sludge discharged to the anaerobic digestion facility as a result of gravity thickening was reduced from 100,000 gpd at 3 percent solids to 66,200 gpd at 4.5 percent solids. The gravity thickener was designed using a mass loading of 10 psf/day. A hydraulic overflow rate of 600 gpd/sq ft was achieved by recycling final effluent to the mixing chamber ahead of the thickener as shown on Figure 10-4.

The capital costs for gravity thickening in this example are estimated at $305,000. These costs include one gravity thickener, a mixing chamber, effluent recycle capacity, and an allowance for appropriate connecting piping. They do not include the allowance for engineering design, bonding, and construction supervision.

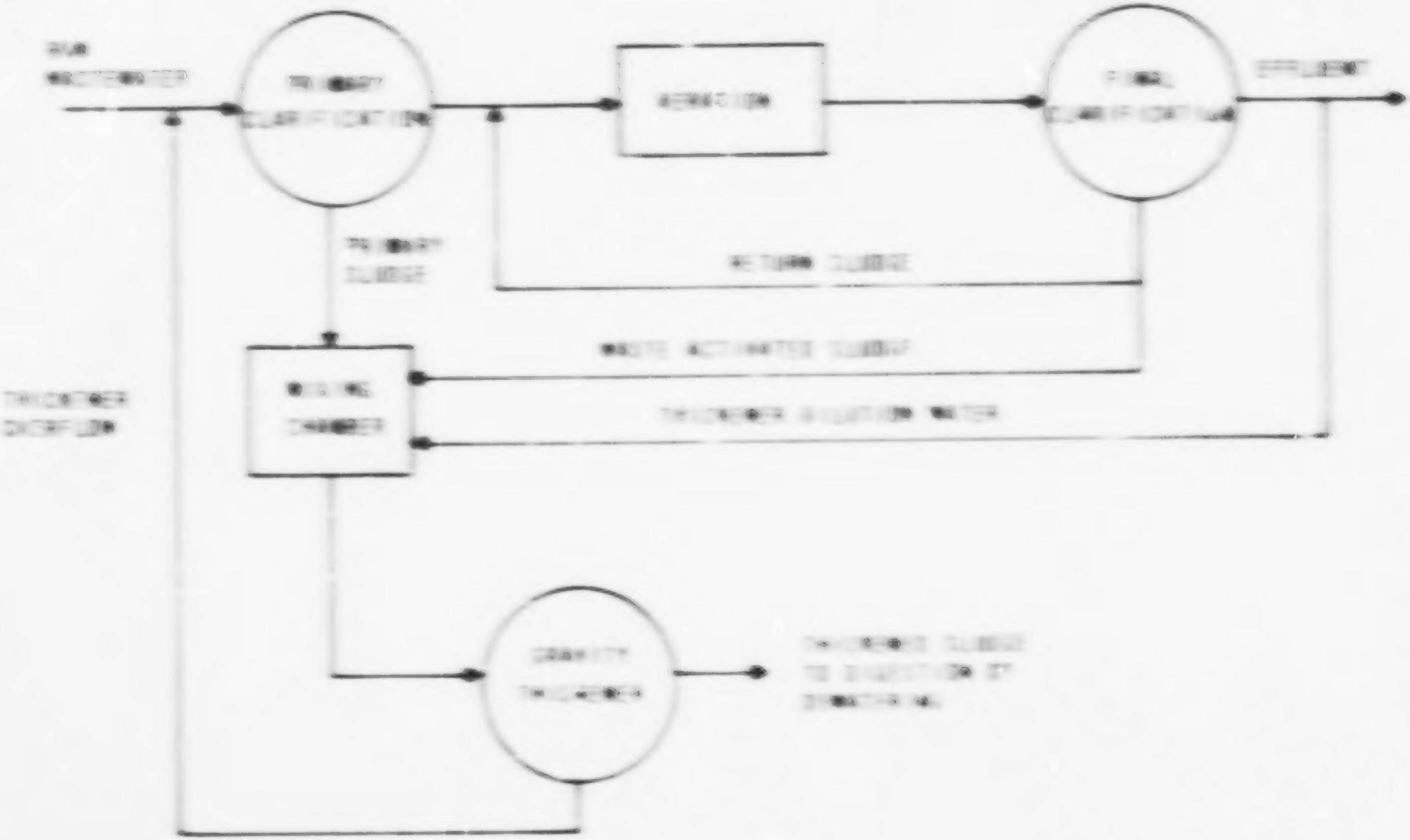
10.4 Air Flotation

The use of air flotation is limited primarily to thickening of sludges prior to dewatering. Used in this way, the efficiency and/or capacity of the subsequent dewatering units can be increased and the volume of supernatant from the subsequent digestion units can be decreased. Existing air flotation thickening units can be upgraded by the optimization of process variables and by the utilization of polyelectrolytes.

Air flotation thickening is best applied to thickening waste activated sludge. With this process it is possible to thicken the sludge to 6 percent, while the maximum concentration attainable by gravity thickening without chemical addition is 2 to 3 percent (1). The air flotation process can also be applied to mixtures of primary and waste activated sludge. The greater the ratio of primary sludge to waste activated sludge, the higher the permissible

FIGURE 10-4

UPGRADING SLUDGE HANDLING FACILITIES USING GRAVITY THICKENING


RAW WASTEWATER
PRIMARY CLARIFICATION
AERATION
FINAL CLARIFICATION
EFFLUENT
PRIMARY SLUDGE
RETURN SLUDGE
WASTE ACTIVATED SLUDGE
THICKENER OVERFLOW
MIXING CHAMBER
THICKENER DILUTION WATER
GRAVITY THICKENER
THICKENED SLUDGE TO DIGESTION OR DEWATERING

solids loading to the flotation unit. Due to the high operating costs, it is generally recommended that air flotation be considered only for thickening waste activated sludge (17).

10.4.1 Process and Design Considerations

The most commonly used type of air flotation unit is the dissolved air pressure flotation unit. A schematic flow diagram for a typical unit is illustrated in Figure 10-5. In this unit, the recycled flow is pressurized from 40 to 70 psig and then saturated with air in the pressure tank. The pressurized effluent is then mixed with the influent sludge and subsequently released into the flotation tank. The excess dissolved air then separates from solution, which is now under atmospheric pressure, and the minute (average diameter 80 microns) rising gas bubbles attach themselves to particles which form the sludge blanket (1). The thickened blanket is skimmed off and pumped to the downstream sludge handling facilities while the subnatant is returned to the plant.

The following table is a summary of typical parameters used in the design of air flotation thickening units.

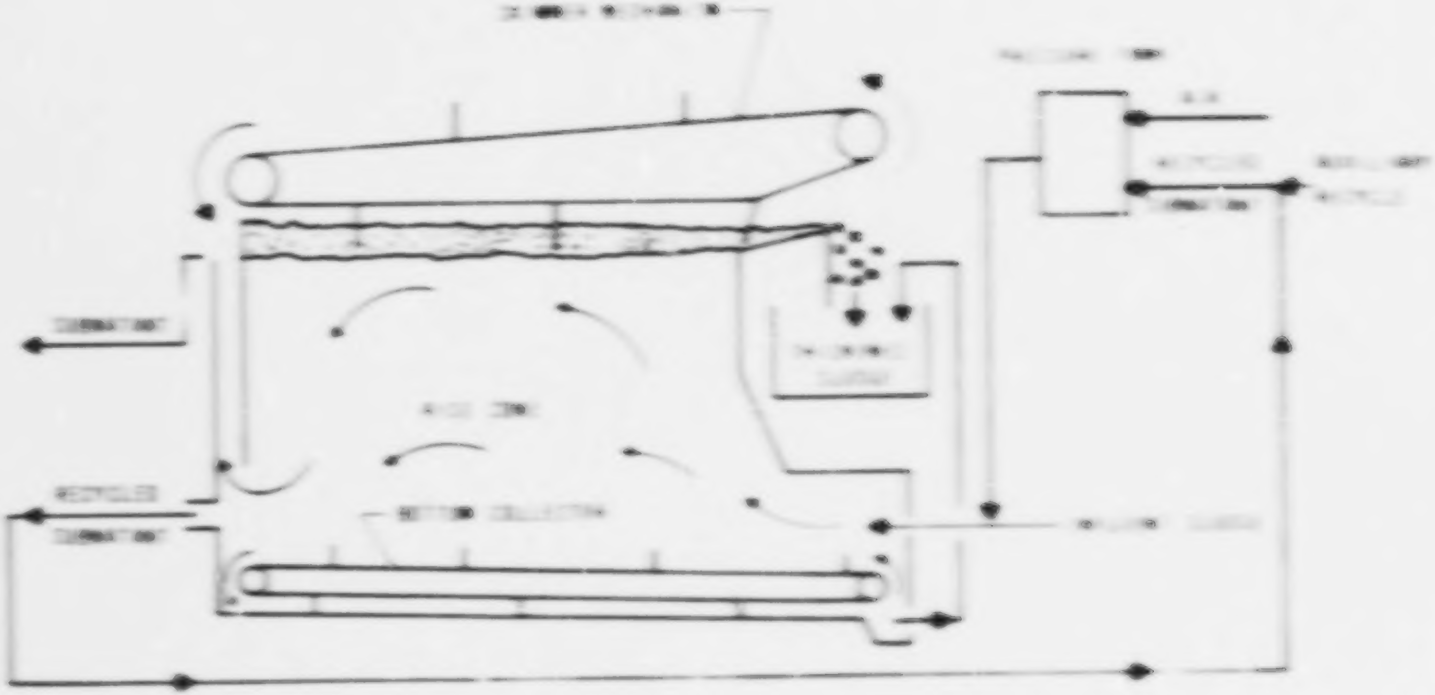
TABLE 10-3
TYPICAL AIR FLOTATION DESIGN PARAMETERS

Parameter	Without Polyelectrolyte	With Polyelectrolyte
Air Pressure, psig	40-70	
Effluent Recycle Ratio, percent of Influent Flow	30-150	
Air to Solids Ratio, lb air/lb solids	0.02	
Solids Loading, psf/day	10 (average)	20-40
Polyelectrolyte Addition, lb/ton dry solids	0	5-10
Solids Capture, percent	70-90	90-98
Total Solids, percent		
Unthickened	0.5-1.5	
Thickened	4-6	

A detailed discussion of the above parameters can be found in references (1) (17) (18) (19).

FIGURE 10-5

SCHEMATIC OF AN AIR FLOTATION UNIT


AIR
BOTTOM COLLECTOR

Bench scale flotation tests have been devised for air flotation design, but poor correlations have generally been obtained with full scale performance (17,19). Therefore, pilot unit results are recommended to determine optimum recycle ratio, chemical requirements, and general applicability of air flotation to sludge thickening.

Typical operating data for various air flotation units are presented in Table 10-4. Combined primary and activated sludge produces a more concentrated float sludge than waste activated sludge alone. Polyelectrolyte and/or chemical addition allows greater solids loading and improves solids recovery without substantially increasing the float solids concentration. However, the increased operating cost of the polyelectrolyte may largely offset the capital cost savings resulting from the reduction of thickener area.

TABLE 10-4
AIR FLOTATION THICKENING PERFORMANCE DATA

Type of Sludge	Solids Loading psf/day	Feed Solids percent	Float Solids percent	Solids Recovery percent	Reference
Waste Activated	1.2 to 10	0.5 to 1.5	4.0 to 6.0	85 to 95	20
Waste Activated*	2.0 to 10	0.5 to 1.5	4.0 to 5.0	93 to 98	21
Waste Activated	4.9	0.84	3.9	85	22
Waste Activated	7.8	0.77	3.7	90	22
Waste Activated	10.0	0.45	4.6	84	22
Waste Activated	10.2	0.80	6.5	91	22
Waste Activated	20.0	0.46	4.0	80	22
Combined Primary and Waste Activated	2.4 to 10	1.5 to 3.0	6.0 to 8.0	85 to 95	20
Combined Primary and Waste Activated	24	0.64	8.6	94	22
Combined Primary and Waste Activated	16.6 10.7	2.00 1.77	7.4 5.4	94 90	23 24

*3 to 6 lb polyelectrolyte/ton dry solids

The following example illustrates the use of prethickening by air flotation prior to anaerobic digestion.

Existing anaerobic digesters were experiencing unstable operation due to the increased volume of sludge produced by an increase in plant flow from 10 to 16 mgd. Operational data from the overloaded and the upgraded plant is presented in Table 10-5. Prior to upgrading, waste activated sludge was recycled to the primary clarifier. The volume of the combined sludge was 100,000 gpd at 3 percent solids. The increased sludge volume due to the plant overloading decreased the detention time in the digesters from 17 to 11.25 days. To improve the operation of the existing digesters, it was necessary to reduce the sludge volume to increase the digester detention time. To reduce sludge volume, thickening of the waste activated sludge by air flotation was considered.

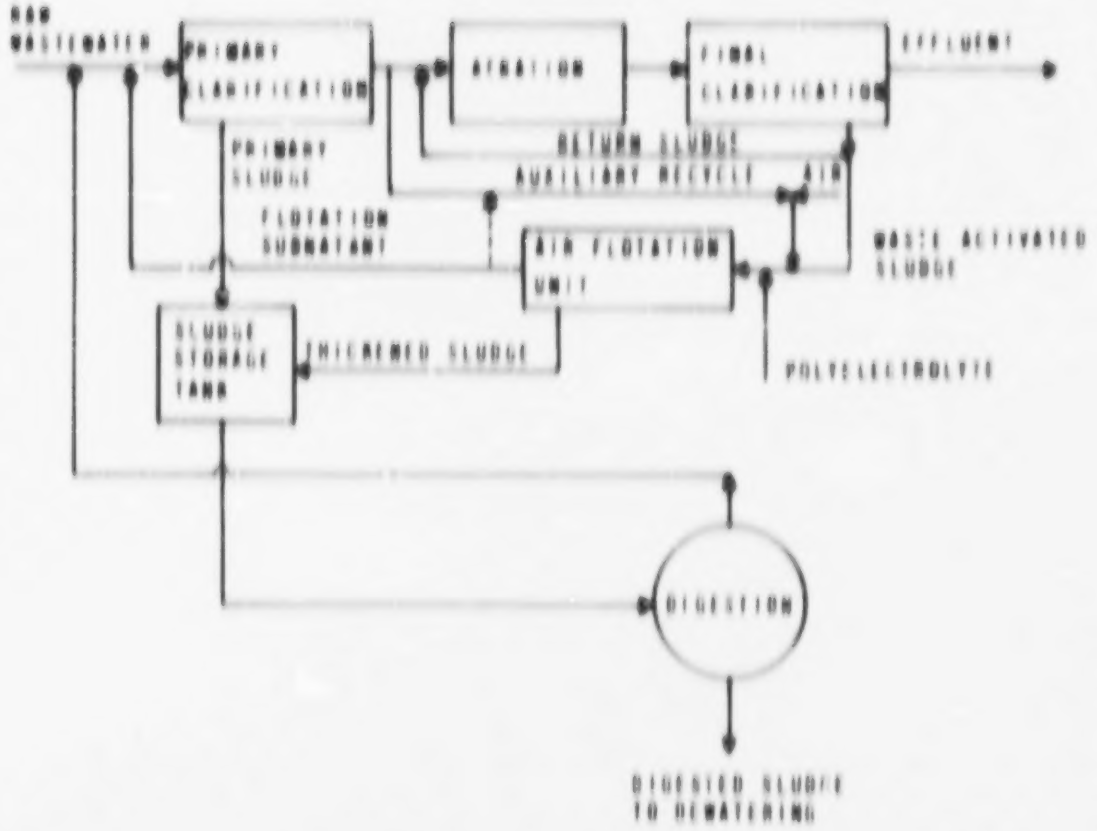
TABLE 10-5
EXAMPLE OF
UPGRADING SLUDGE HANDLING FACILITIES
USING AIR FLOTATION THICKENING

Description	Overloaded Plant	Upgraded Plant
Primary Solids, lb	15,300	15,300
Primary Sludge Volume, gpd		30,600 (6%)
Waste Activated Solids, lb	9,700	9,700
Waste Activated Sludge Volume, gpd		29,100 (4%)
Combined Solids, lb	25,000	25,000
Combined Sludge Volume, gpd	100,000 (3%)	59,700 (5%)
Digester Hydraulic Detention Time, days	11.25	18.8

The flow diagram for the upgraded plant is shown on Figure 10-6. As a result of separate thickening of the waste activated sludge, it is expected that the primary sludge can be concentrated to 6 percent in the primary clarifiers. The total volume of sludge discharged to the anaerobic digestion facility due to the separate thickening of the waste activated sludge and the improved solids concentrations in the primary clarifier is expected to be 59,700 gpd, compared to 100,000 gpd prior to upgrading. The air flotation system was designed using an air pressure of 50 psig, an effluent recycle of 100 percent, and a solids

FIGURE 10.6

UPGRADING SLUDGE HANDLING FACILITIES USING AIR FLOTATION THICKENING


RAW WASTEWATER
PRIMARY CLARIFICATION
AERATION
FINAL CLARIFICATION
EFFLUENT
PRIMARY SLUDGE
RETURN SLUDGE
AUXILIARY RECYCLE
AIR
FLOTATION SUBNATANT
AIR FLOTATION UNIT
WASTE ACTIVATED SLUDGE
SLUDGE STORAGE TANK
THICKENED SLUDGE
POLYELECTROLYTE
DIGESTION
DIGESTED SLUDGE TO DEWATERING

loading of 25 psf/day. It is anticipated that the polyelectrolyte dosage requirements will be 5 lb/ton of dry solids.

The capital costs for air flotation thickening in this example are estimated at $181,000. These costs include two air flotation units, a polyelectrolyte addition system, and appropriate connecting piping. They do not include an allowance for engineering design, bonding, and construction supervision.

Where separate thickening of primary and waste activated sludge is indicated, separate gravity thickening of primary sludge in conjunction with air flotation thickening of waste activated sludge will normally provide minimum thickening tank area. The following example has been prepared to demonstrate this concept.

	Combined Gravity Thickening		Separate Thickening
Plant Flow, mgd		50	
Primary Sludge, lb/day		110,000	
Waste Activated Sludge, lb/day		40,000	
Solids Loadings, psf/day			
Combined Gravity Thickening	12		
Primary Sludge Thickening			22
Waste Activated Sludge Thickening			25
Area Required, sq ft			
Combined Thickening	12,500		
Primary Sludge Thickening			5,000
Waste Activated Thickening			1,600
Total	12,500		6,600

10.5 Centrifugation

The centrifugation process has been successfully used for many years in industry for separating liquids of different density, for thickening slurries, and for removing solids. Although the potential value of centrifuges for wastewater treatment has been recognized for quite some time, only recently have they been installed for regular use in wastewater treatment plants. The increasing use of centrifuges in the wastewater treatment field is the result of recent improvements in centrifuge design, the availability of reliable performance data, and the advantages of centrifuges in certain instances over other sludge processing facilities.

This discussion is limited to performance data for centrifuges used in sludge thickening and the applicability of centrifuges for upgrading sludge thickening facilities. The types of centrifuges used for sludge thickening and dewatering are discussed in Section 12.4.1 and illustrated on Figure 12-2. The centrifuge process and design principles discussed in Sections 12.4.2 and 12.4.3 are applicable to both sludge thickening and sludge dewatering applications.

10.5.1 Centrifuge Performance in Sludge Thickening

Due to high maintenance and power costs, centrifugal thickening is generally used only when space limitations or sludge characteristics make other methods unsuitable. Performance data for centrifuges used to thicken various types of sludge are shown in Table 10-6.

A disc centrifuge can thicken waste activated sludge to 4 to 5 percent solids without polyelectrolytes. However, it is essential that the unit be preceded by coarse and fine screening plus cyclonic grit removal to prevent clogging of the discharge nozzles.

A basket centrifuge can also thicken waste activated sludge. It can attain 8 to 10 percent discharge solids, but thickens at a much lower rate than a disc centrifuge. No sludge prescreening is required with the basket centrifuge.

A solid bowl centrifuge can thicken waste activated sludge to 5 to 8 percent solids. The addition of polyelectrolytes to the feed sludge significantly improves the solids feed rate and/or capture efficiency in this type unit.

Centrifugal thickening of waste activated sludge was extensively studied at the Chicago Sanitary District. Fitch and Kennedy (16) evaluated both disc type and solid bowl centrifuges. The disc type machine concentrated the activated sludge to about 7.0 percent at 6,000 rpm, but operational problems made its use impractical. Clogging of the sludge discharge nozzles required repeated maintenance. Solid bowl centrifuges, processing activated sludge alone, thickened sludge from 6.6 to 7.5 percent.

A disc centrifuge has been successfully field tested for thickening waste activated sludge at an eastern Pennsylvania community (26). Using a 30-inch centrifuge with a 150-hp motor and 300-gpm feed rate, the disc centrifuge produced a 5 percent underflow with 90 percent solids recovery. Since the plant did not have primary treatment, it was necessary to install a screening device ahead of the centrifuge. The screening effectiveness was demonstrated in that the nozzles of the centrifuge did not plug. The combination of effective screening and patented recirculating system (allowing a larger nozzle size) was instrumental to the good performance. Normally, however, disc centrifuges are not recommended where activated sludge treatment has not been preceded by primary treatment.

TABLE 10-6

CENTRIFUGAL THICKENING PERFORMANCE DATA

Type of Sludge	Centrifuge Type	Capacity	Feed Solids	Underflow Solids	Solids Recovery	Polymer Requirement	Reference
		gpm	percent	percent	percent	lb/ton	
Waste Activated	Disc	150	0.75-1.0	5.0-5.5	90+	None	21
Waste Activated	Disc	400		4.0	80	None	21
Waste Activated (after Roughing Filter)	Disc	50-80	0.7	5-7	93-87	None	23
Waste Activated (after Roughing Filter)	Disc	60-270	0.7	6.1	97-80	None	23
Waste Activated	Basket	33-70	0.7	9-10	90-70	None	23
Waste Activated	Solid Bowl	10-12	1.5	9-13	90		24
Waste Activated	Solid Bowl	75-100	0.44-0.78	5-7	90-80	None	25
Waste Activated	Solid Bowl	110-160	0.5-0.7	5-8	65	None	25
					85	< 5	
					90	5-10	
					95	10-15	

10.5.2 Use of Centrifugal Thickening for Upgrading Existing Sludge Handling Facilities

The performance of centrifuges in various applications clearly indicates that centrifugal thickening should be considered when the upgrading of solids handling facilities is required. Centrifuges are a flexible upgrading device because of their applicability in the thickening of various mixtures of sludges.

When used as a thickening device, a centrifuge can upgrade an overloaded anaerobic digester by reducing the volume of feed sludge, thereby increasing digester detention time. In addition, centrifuges can also be used to supplement existing overloaded gravity thickeners.

10.5.3 Process Design and Cost Estimates

The following example is given to illustrate the upgrading of sludge handling facilities through the use of centrifuges.

Existing anaerobic digestion facilities at an activated sludge plant became overloaded due to an increase in plant flow from 10 to 16 mgd. As a result, the detention time in the digesters was reduced to 11.25 days, thereby causing unstable operation of the digesters. Operational data from the overloaded and the upgraded plant are presented in Table 10-7. Prior to upgrading, waste activated sludge was recycled to the primary clarifier.

TABLE 10-7
EXAMPLE OF
UPGRADING SLUDGE HANDLING FACILITIES
USING CENTRIFUGAL THICKENING

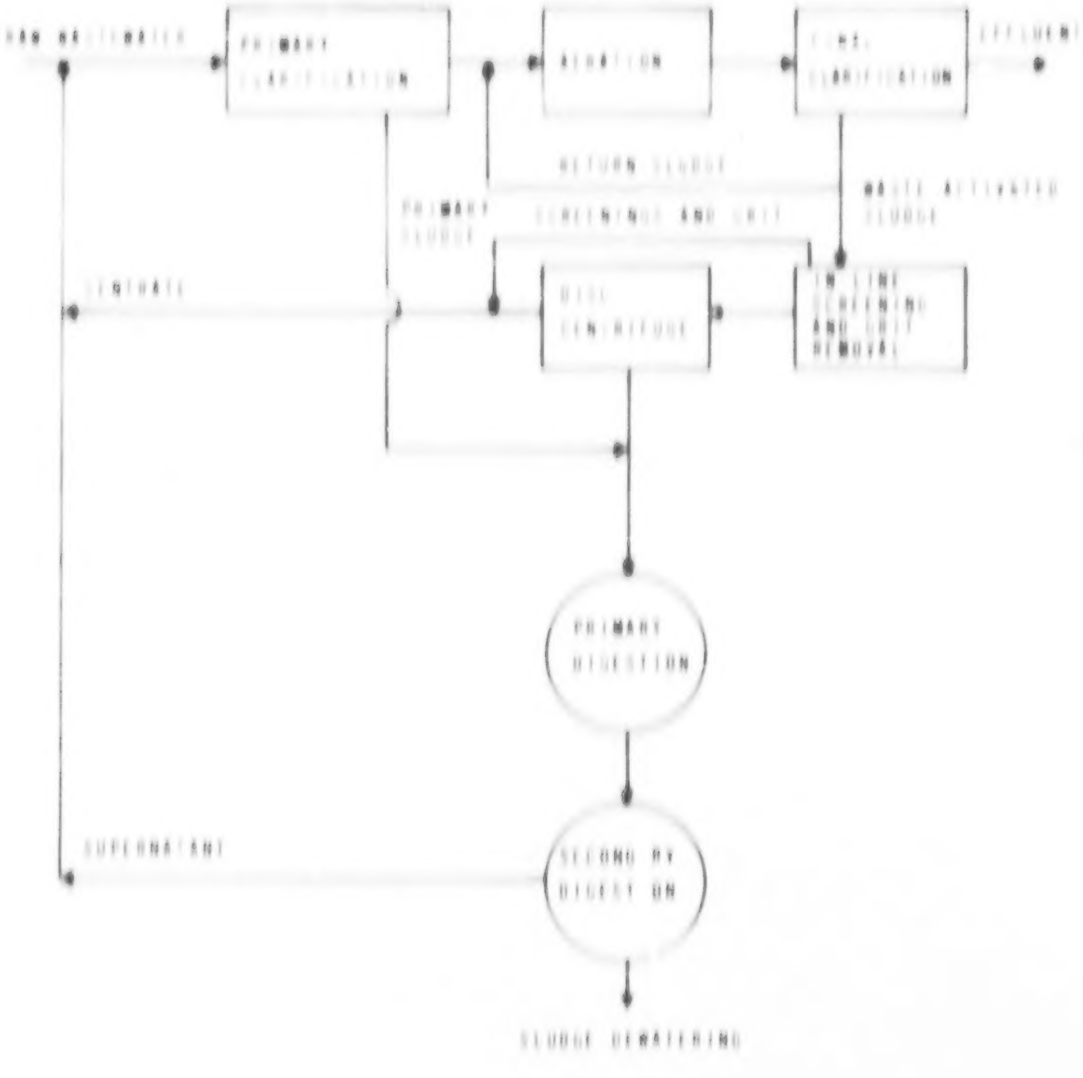
Description	Overloaded Plant	Upgraded Plant
Primary Solids, lb	15,300	15,300
Primary Sludge Volume, gpd		30,600 (6%)
Waste Activated Solids, lb	9,700	9,700
Waste Activated Sludge Volume, gpd		23,300 (5%)
Combined Solids, lb	25,000	25,000
Combined Sludge Volume, gpd	100,000 (3%)	53,900 (5.6%)
Digester Hydraulic Detention Time, days	11.25	21

To upgrade the digestion facilities, it was decided to thicken the waste activated sludge by using a disc centrifuge, as shown on Figure 10-7. It was decided that polyelectrolyte addition was not necessary. The volume of the waste activated sludge, 69,400 gpd at 1.7 percent solids (9,700 lb dry solids/day) before centrifugal thickening was reduced to 23,300 gpd at 5 percent solids. Combination of the thickened waste activated sludge with 30,600 gpd of 6 percent primary sludge resulted in an increase in digester detention time to approximately 21 days.

The capital cost for this upgrading procedure was estimated at $253,000. This cost includes one operating and one standby disc centrifuge, in-line screens, sludge pumps, and appurtenances, but does not include an allowance for engineering, design, bonding, and construction supervision.

FIGURE 10.7

UPGRADING SLUDGE HANDLING USING CENTRIFUGAL THICKENING


RAW WASTEWATER
PRIMARY CLARIFICATION
AERATION
FINAL CLARIFICATION
EFFLUENT
RETURN SLUDGE
WASTE ACTIVATED SLUDGE
PRIMARY SLUDGE
SCREENINGS AND GRIT
CENTRATE
DISC CENTRIFUGE
IN-LINE SCREENING AND GRIT REMOVAL
PRIMARY DIGESTION
SUPERNATANT
SECONDARY DIGESTION
SLUDGE DEWATERING

10.6 References

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CHAPTER 11

SLUDGE STABILIZATION

11.1 General

Sludge stabilization processes are used to convert raw wastewater sludges to inoffensive forms by decreasing the organic content in the sludges or by otherwise rendering them inert. This is especially essential when sludge is to be disposed of on land. The four major sludge stabilization processes are anaerobic digestion, aerobic digestion, heat treatment and lime stabilization. Principal design parameters and upgrading techniques for these processes are discussed in this chapter. Additional information on the design of these processes is available in the Process Design Manual for Sludge Treatment and Disposal (1).

A sludge stabilization process must be viewed as an integral component of the overall sludge handling and treatment system. The selection and design of each of the components within this system are interdependent. Important considerations in sludge processing interrelationships are discussed in Section 10.1.2.

11.2 Anaerobic Digestion

Anaerobic digestion is one of the more commonly employed processes for sludge stabilization in treatment plants over 1 mgd. The process biologically reduces the amount of VSS that must be handled by subsequent dewatering and ultimate disposal operations, renders the organic material nonputrescible and destroys a large number of pathogenic organisms. However, since anaerobic digestion results in a breakdown of coarse solids into finer particles, digested sludges are generally more difficult to dewater than undigested sludges. Also, the high alkalinity of digested sludges may require a greater dosage of conditioning chemicals than would be required with undigested sludges to achieve comparable performance in the dewatering process.

The major gaseous end product of anaerobic digestion is methane, which is often used as a source of fuel in wastewater treatment plants. The digested sludge is an excellent soil conditioner and has found some utility for this purpose.

A major advantage associated with anaerobic digestion is its low energy requirement. Power consumption is much less than that required for aerobic stabilization or heat treatment. The methane produced is normally more than sufficient to generate the heat required to maintain optimum digester temperatures and thus provides a surplus energy source that can be used elsewhere in a wastewater treatment plant.

11.2.1 Process Biochemistry

Anaerobic digestion of sludge is a complex biochemical process employing several groups of anaerobic and facultative organisms. In general, the process can be considered to consist of two steps. In the first step, facultative organisms called "acid formers" degrade the complex organics of wastewater sludge to volatile organic acids, primarily acetic acid. In the second step, these volatile acids are fermented to methane and carbon dioxide by a group of strict anaerobes called "methane bacteria."

The more important of these two phases is the methane fermentation phase because:

1. The only mechanism of COD or BOD removal is the production of methane. Acid production only solubilizes the complex organics; it does not accomplish stabilization.

2. This step has been found to be the rate limiting step in the reaction sequence.

The primary reason why the methane fermentation step is rate limiting is that the reproduction rate for these organisms is quite low relative to that of other groups of bacteria. For example, the doubling time of the acid formers is several hours while that of the methane formers is, under ideal conditions, four days. Thus, even if a temporary difficulty in the system arose, it would be much harder for the methane organisms to adjust than for the acid formers. In addition, it has been found that the environmental conditions required to maintain optimum performance by the methane organisms are much more restrictive than for the acid-forming organisms. Consequently, most of the effort in upgrading existing digesters should be expended to insure that the methane fermentation step is carried out as efficiently as possible.

Improperly functioning digestion systems can be upgraded by applying procedures which will make the systems more closely approach optimum performance conditions. Therefore, before upgrading existing digesters, the environmental factors discussed below should be investigated to determine if the units are operating at their full potential.

11.2.2 Environmental Conditions for Optimum Performance

Knowledge of the range of environmental conditions which favor optimum performance of anaerobic digestion is not as extensive as desired. A summary of the state of knowledge is given below.

11.2.2.1 pH

Close pH control is required for this process because methane bacteria are extremely sensitive to slight changes in pH. The usual pH range required is 6.6 to 7.4. In general, it is wise to maintain the pH as close to 7.0 as possible.

In a properly operating anaerobic digester, pH is maintained naturally by a bicarbonate buffer system due to the great quantity of carbon dioxide produced during methane fermentation. The pH is a function of the bicarbonate alkalinity of the digesting liquor and the fraction of CO_2 in the digester gas. Figure 11-1 prepared by McCarty (2) illustrates this relationship. Because of the significance of pH control in digester operation, it is most important that the dynamic nature of buffer destruction and formation in the digester be understood. This process is reviewed in the following equations for simple carbohydrates such as glucose. The equations mentioned are equally applicable to digestion of sludge.

$$C_6H_{12}O_6 \xrightarrow{\text{acid formers}} 3CH_3COOH$$

$$3CH_3COOH + 3NH_4HCO_3 \longrightarrow 3CH_3COONH_4 + 3H_2O + 3CO_2$$

$$3CH_3COONH_4 + 3H_2O \xrightarrow{\text{methane bacteria}} 3CH_4 + 3NH_4HCO_3$$

The first equation represents the breakdown of glucose to acetic acid by acid-forming bacteria. The acid is neutralized, as shown in the second equation, by the bicarbonate buffer. If sufficient buffer is not present, the pH would drop, and the conversion of acetate to methane, as shown in the third equation, would be inhibited. During the third reaction, the buffer consumed in the second reaction is reformed.

A dynamic equilibrium between buffer formation and destruction is maintained when the process is proceeding satisfactorily. However, when an upset occurs, it is usually the methane bacteria rather than the acid formers which are adversely affected. Therefore, net buffer consumption takes place, and the process is in danger of pH failure. When this occurs, an external source of alkalinity such as lime must be added to maintain the pH in the proper range.

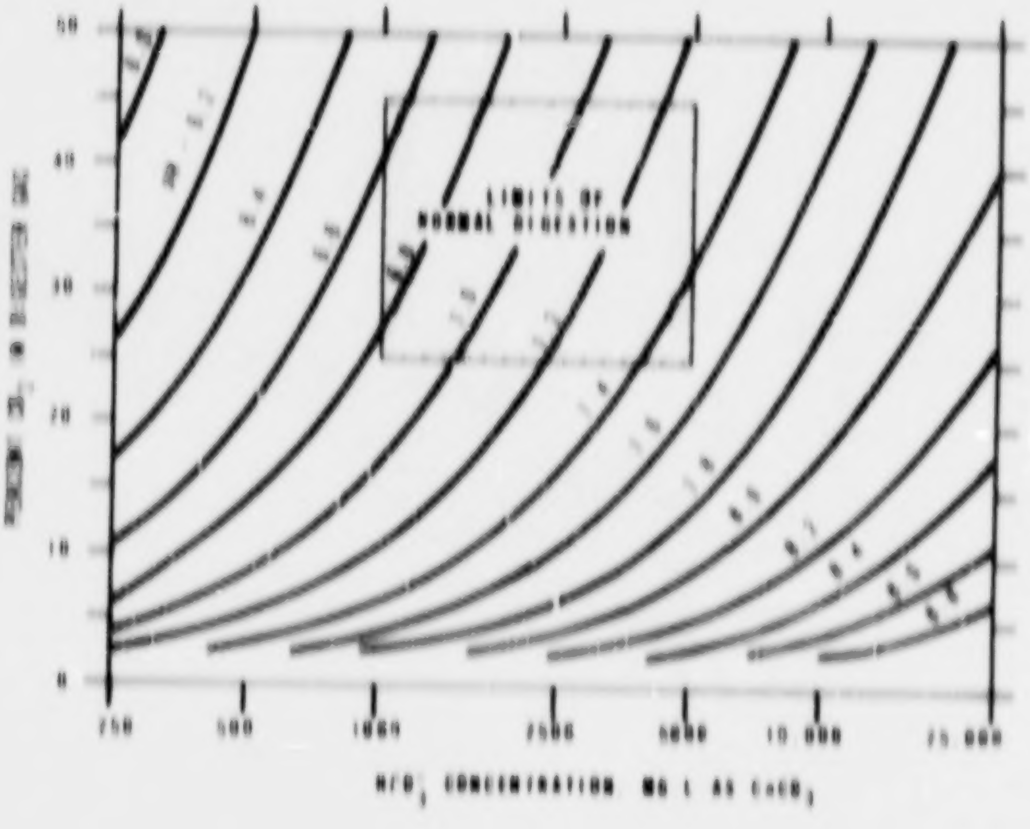
Figure 11-1 indicates that the bicarbonate alkalinity should be maintained at a minimum level of 1,000 mg/l as $CaCO_3$ to ensure adequate pH control. To determine the bicarbonate alkalinity, both the volatile acid concentration and the total alkalinity must be measured. Then,

$$\text{Bicarbonate Alkalinity} = \text{Total Alkalinity} - 0.8\ \text{Volatile Acids}$$

The 0.8 factor in the above equation is required to convert the volatile acid units from mg/l as acetic acid to mg/l as $CaCO_3$, the equivalent alkalinity unit.

FIGURE 11-1

RELATIONSHIP BETWEEN pH AND BICARBONATE CONCENTRATION



It should be noted that in the second and third equations, ammonium bicarbonate was used as the form of the alkalinity. This represents the situation in wastewater sludge digestion where large concentrations of ammonium result from the destruction of protein. In fact, the maximum value of the total alkalinity is set by the concentration of the ammonium ion. The carbon dioxide generated in the methane fermentation will not form negatively charged bicarbonate (the buffer) unless an equivalent quantity of cation is present. This is provided by the destruction of natural protein with the release of positively charged ammonium. If a cation is not present to force formation of bicarbonate buffer, self-regulation of pH in the digestion process is not possible. In this case, alkaline material must be added continuously to control the pH. For example, the anaerobic degradation of glucose, illustrated in the first three equations, would require the addition of an external source of alkalinity. It is not necessary that ammonium bicarbonate be utilized for this purpose; in fact, in terms of cost and avoidance of potential toxicity, another bicarbonate salt might be favored.

In general, this difficulty will not be experienced in wastewater sludge digestion unless either a high carbohydrate fraction from an industrial waste is present in the sludge or a very thin sludge is being treated.

11.2.2.2 Temperature

The temperature response of methane bacteria is the same as other bacterial groups. Although thermophilic methane bacteria exist, it is generally not economically feasible to heat sludge to this temperature range. Thus, digestion of wastewater sludge is conducted in the mesophilic range. The optimum temperature in this range is 35 deg C (95 deg F). More important than maintenance of a particular temperature is maintenance of the chosen temperature at a constant level. A temperature change of 1 or 2 deg C is sufficient to disturb the dynamic balance between the acid formers and the methane formers. This will lead to an upset because the acid formers will respond much more rapidly to changes in temperature than will the methane bacteria. When heat exchangers are used to upgrade the performance of existing digesters, care must be exercised to avoid wide temperature variations and excessive temperature in the heat exchanger itself.

11.2.2.3 Nutrients

Speece and McCarty (3) have done the most definitive work on the macro-nutrient and micro-nutrient requirements of methane bacteria. As these authors indicate, domestic wastewater appears to contain all of the nutrients required by these organisms. However, due to the uncertainty of the precise nutritional requirements of methane bacteria, difficulty may be encountered in digestion when a considerable fraction of the wastewater is of industrial origin.

11.2.2.4 Toxic Materials

Substances which may be present in municipal sludges in concentration ranges which can produce toxicity include heavy metals, sulfides, surface-active agents, light metals, and certain organics. All of these can gain entrance to wastewater sludge from industrial sources. In addition, light metal cations will enter sludge if an alkaline material is added to control the pH. Several papers (4) (5) (6) review the best engineering data available on toxicity. Reference should be made to these papers for complete information. General information on some substances is shown in Table 11-1.

TABLE 11-1
CONCENTRATIONS WHICH WILL CAUSE A TOXIC SITUATION
IN ANAEROBIC DIGESTION OF MUNICIPAL SLUDGES

Substance	Concentration mg/l
Sulfides	200
Heavy Metals[1]	> 1
Sodium	5,000 - 8,000
Potassium	4,000 - 10,000
Calcium	2,000 - 6,000
Magnesium	1,200 - 3,500
Ammonium	1,700 - 4,000
Free Ammonia	150

[1] Soluble.

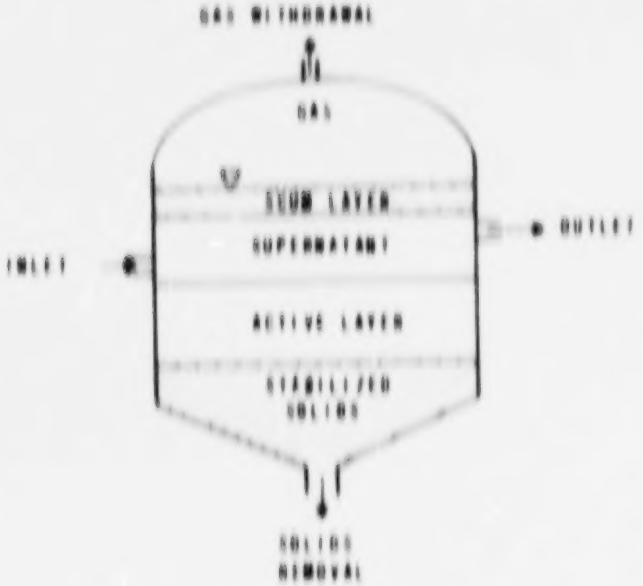
It must be emphasized that the values in this table are only guides. If toxicity is suspected, a thorough analysis of all the chemical constituents of the sludge must be made before definite conclusions can be drawn. Potential solutions to toxicity problems, other than elimination from the wastewater, should be evaluated in small-scale digesters of the type used in laboratory investigations.

11.2.3 Anaerobic Digestion Systems

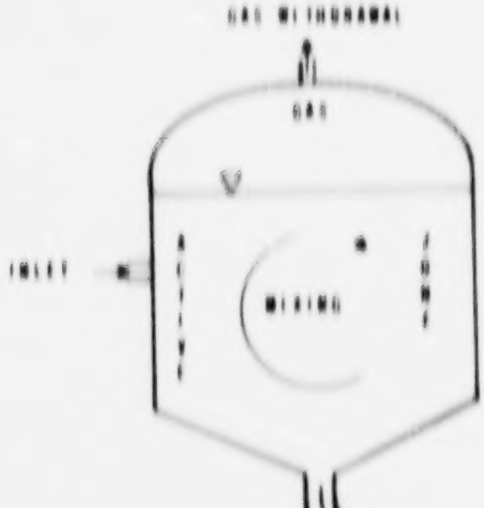
Prior to a discussion of procedures for upgrading the performance of anaerobic digestion systems, a description of typical anaerobic digestion systems will be presented. Figure 11-2 illustrates the two types of digestion systems in use at present.

FIGURE 11-2

ANAEROBIC DIGESTION SYSTEMS


GAS WITHDRAWAL
GAS
SCUM LAYER
SUPERNATANT
OUTLET
INLET
ACTIVE LAYER
STABILIZED
SOLIDS
SOLIDS
REMOVAL

LOW RATE DIGESTER


GAS WITHDRAWAL
GAS
INLET
ACTIVE
MIXING
ZONE
OUTLET

HIGH RATE DIGESTER

In a low rate system, the tank is not mixed and, in some cases, is not heated. Sludge is added at the top and withdrawn at the bottom. Stratification develops in the system due to a lack of mixing. In general, this can be classified as a plug flow system. Because of the lack of mixing and consequent stratification, much of the digester volume is wasted, and many operational problems result. In this type of digester, acidification sometimes takes place in the top and middle layers while methane fermentation is confined to the lower layers. This can lead to areas of low and high pH in the system, which restrict optimum biological activity. Also, chemicals added for pH control are not dispersed throughout the tank, and their effectiveness is limited. Grease breakdown is poor because the grease tends to float to the top of the digester while the methane bacteria are confined to the lower levels. Methane bacteria are removed with the digested sludge and are not recycled to the top, where they are required. During progression from top to bottom of the digestion tank, the sludge is compressed and gradually dewatered.

The high rate system differs from the low rate system in that the contents are well mixed, either continuously or intermittently, and the digester is heated. This procedure avoids most of the difficulties inherent in low rate systems. Consequently, this system operates well at lower HRT values and higher organic loading rates. Table 11-2 compares design criteria for low rate and high rate digesters.

TABLE 11-2
TYPICAL DESIGN CRITERIA FOR LOW RATE AND HIGH RATE ANAEROBIC DIGESTERS

Parameter	Low Rate	High Rate
Hydraulic Retention Time (HRT), days	30 to 60	15 to 20
VSS Solids Loading, pcf/day	0.04 to 0.1	0.15 to 0.40
Volume Criteria, cu ft/capita		
Primary Sludge	2 to 3	1 1/3 to 2
Primary Sludge + Trickling Filter Sludge	4 to 5	2 2/3 to 3 1/3
Primary Sludge + Waste Activated Sludge	4 to 6	2 2/3 to 4

Various mixing systems have been successfully utilized in digesters. These include:

1. Single or multiple draft tubes through which sludge is circulated by a turbine mixer set in the tube.

2. Digester gas recirculation through diffusers in the base of the digester or drop pipes.

Specific design information for these systems can be obtained from the various manufacturers.

Sludge heating is accomplished either (1) by circulating hot water through coils mounted on the inner wall of the digestion tank, (2) by a heat exchanger furnished as part of the mixer draft tube or (3) by circulating sludge through an external heat exchanger. The latter method is preferred since it has been found that the coils inside the digestion tank are easily caked with partially dried sludge. The sludge circulation rate to the external heat exchanger depends on the efficiency of the heat exchanger and the temperature differential. Design of a high-rate digestion system must include a heat balance to determine fuel requirements. The WPCF Manual of Practice No. 8 (7) presents in detail the procedure for making such a heat balance. This discussion includes valuable data on the fuel value of sludge gas and the insulation characteristics of typical digester construction materials.

One difficulty with high-rate digestion is that the sludge leaving the digester is thinner than the incoming sludge due to solids destruction. Secondary digesters are normally added to many high-rate digestion systems to concentrate the sludge and provide supernatant separation. These units are designed primarily as settling tanks, but also provide a source of seed sludge in case of digester upset, and sludge storage where intermittent withdrawals for dewatering or drying are made. If suitably equipped, they can also serve as standby digesters.

11.2.4 Upgrading Existing Anaerobic Digestion Facilities

The first step in upgrading a digester is to evaluate process performance. This can be accomplished with the aid of such tests as volatile acids and alkalinity and by a digester gas analysis. Any sudden rise of volatile acids indicates that the system is out of biochemical balance. A rise in the CO_2 fraction in the gas or a decrease in methane production per pound of volatile solids added will also indicate upset. However, the volatile acid test is more sensitive. When an upset occurs, an alkaline material must be added to maintain the bicarbonate alkalinity above 1,000 mg/l as $CaCO_3$. An easily soluble bicarbonate salt, such as $NaHCO_3$, is best for this purpose. Care must be exercised not to exceed the level at which the cation of the alkaline material will cause toxicity. If this is a potential problem, a mixture of alkaline salts should be used. Kugelman and McCarty (5) have described methods of preventing cation toxicity by adding appropriate quantities of cation antagonists.

Control of pH during an upset is only a temporary measure. The cause of the upset must be located and eliminated. Sometimes this is easy. For example, heavy metal toxicity can be completely eliminated by precipitation of the metal in the digester as the sulfide (8). In other cases, only exclusion of the toxin from the system will suffice.

The major upgrading technique for low rate digesters is conversion to high rate digestion. To maintain high rate digestion, the following conditions are necessary:

1. Solids thickening to increase volatile solids loading and HRT
2. Complete mixing of digester contents
3. Uniform solids feed and withdrawal
4. Temperature control system capable of maintaining a uniform temperature at a point within the range of 30 to 35 deg C
5. An average HRT of at least 15 days.

The principal techniques used for upgrading high-rate digesters are to increase feed solids concentration, to provide a secondary digester for liquid-solids separation and to increase HRT.

The relationship between thickening of solids, HRT and solids loading has been illustrated by Sawyer (9) and is shown on Figure 11-3. This relationship points out the importance of thickening the solids prior to digestion. Sludge thickening is discussed in Chapter 10. Sections 10.3.5, 10.4.2 and 10.5.3 illustrate how the solids loading capacity of anaerobic digesters can be increased by implementing sludge thickening techniques.

There are restrictions on the degree to which raw sludge can be thickened. These include the difficulty of pumping thick sludge and the maintenance of adequate mixing in the digester. Generally, sludge need not be thickened beyond about eight percent solids to optimize digester volume utilization. If it is desired to thicken beyond this point, adequate studies of the sludge flow characteristics must be made to evaluate pumping and mixing requirements.

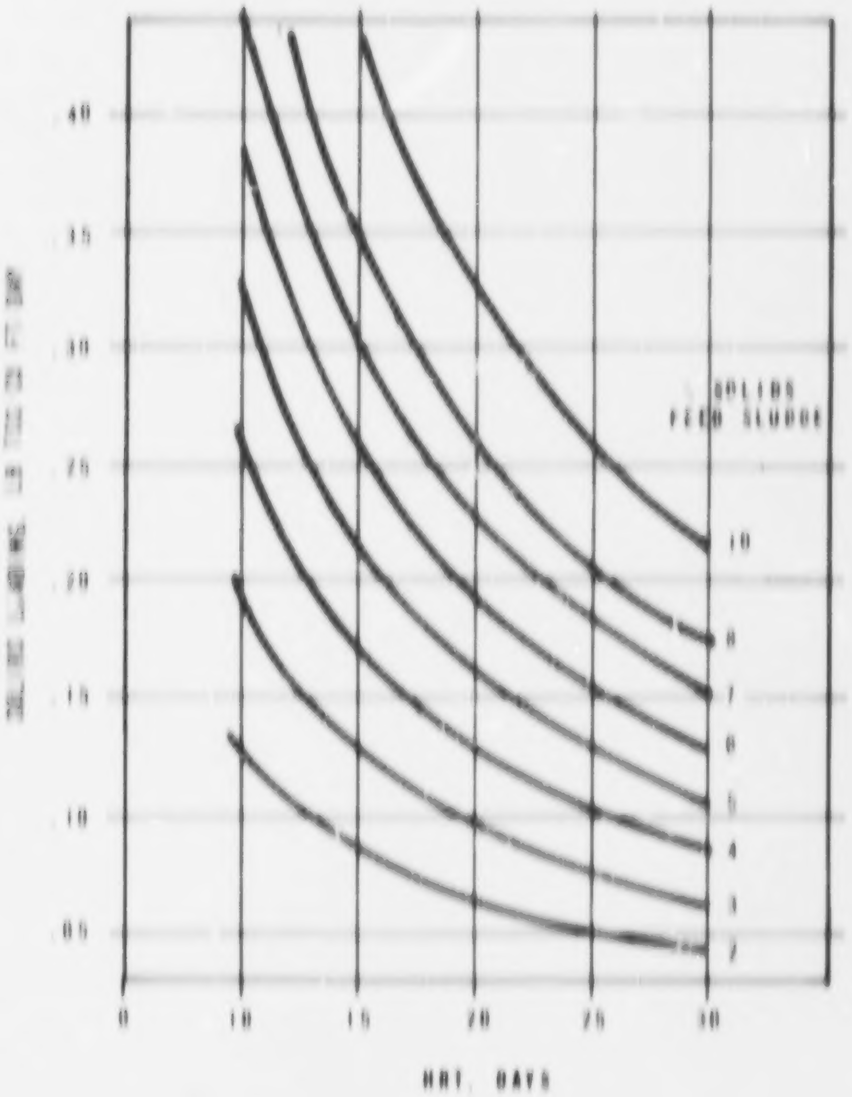
The addition of digester mixing by itself can have a significant beneficial effect on digester performance. At the City of Pontiac, Michigan (10), modifications were made to an existing digester by adding a gas recirculation unit to improve mixing. Mixing inhibited scum formation, improved heat transfer and provided a more stable digestion process. Many other communities have had similar success (11).

Experiences at Chicago (12) have shown that digesters can be upgraded to operate at VSS loadings of 0.2 pcf/day and at an HRT of 10 days. Complete mixing is necessary to operate under these loading conditions.

Bergen County, New Jersey, is the best reported example of upgrading low rate digesters to high rate digesters (13). In 1951, a 30 mgd activated sludge plant was constructed with four 1.3 million gallon digesters. In 1961, the capacity of the plant was increased to 50 mgd, and

FIGURE 11-3

RELATIONSHIPS BETWEEN SLUDGE SOLIDS DIGESTER LOADINGS, AND DETENTION TIME (9)



the existing digesters were modified to high rate. In 1969, additional studies were conducted on a full-scale basis, and it was found that two of the original digesters could handle the entire loading from the existing 50 mgd activated sludge plant. On this basis, it has been projected that the original four digesters will be able to handle the increased solids loading when the plant is expanded to 75 mgd. It is significant to note that even at these higher loadings, the anaerobic process was very stable and the efficiency of the process remained the same.

Successful upgrading of existing digestion facilities at Bergen County was accomplished by the following methods:

1. Completely mixing the digester contents.

2. Prethickening the primary and secondary sludge to an average concentration of 6.3 percent (range 5.2 to 7.5 percent).

3. Increasing the applied solids loading from 0.22 pcf/day to 0.5 pcf/day as a result of thickening.

4. Decreasing the HRT from 22 to 10 days.

The loading to two-stage high-rate digestion systems may be increased by recirculating the concentrated digested sludge from the secondary stage back to the primary stage, since this effectively increases the SRT at the same digester HRT. As with activated sludge systems, the limiting factor is the solids-liquid phase separation. Perhaps a degasification process can be included between the primary and secondary tanks to aid in the separation, as is done with the anaerobic contact process (14). When satisfactory phase separation is obtained in the secondary digester, adequate SRT's can be maintained while decreasing HRT's to less than 10 days (15). A more stable digestion process then results, due to higher populations of methane bacteria in the primary stage and to the lessening of toxicity effects at longer SRT's.

Digesters can also be upgraded by recirculating a portion of the digested sludge back to the thickening units and mixing it with the incoming combined sludge and effluent recycle water. This procedure has been reported by Torpey and Melbinger (16) with modified aeration plants in New York City, and was originally adopted to improve the pumping characteristics of highly concentrated digester feed sludge of 10 to 14 percent solids. It appears that the thickening process was improved because the digested sludge was incorporated into the pore spaces of the mixed primary and waste activated sludge, thereby eliminating the typical gel structure produced by grease in the raw sludge. It was also found that recycling digested sludge improved digester performance due to the seeding of the combined sludge prior to digestion and because of the greater SRT thereby afforded. Volatile solids reduction was also increased. A digested sludge recycle of 50 percent appeared to be optimum for the New York City plants, and a net volume reduction of digested sludge from 197 cu ft/million gallons to 112 cu ft/million gallons was achieved.

It must be pointed out that conversion to high-rate digestion is not a cure-all, especially if digested sludge is to be dewatered prior to final disposal. Methane production and volatile solids reductions are approximately the same at high rate as at low rate, but indications are that supernatant separation and dewatering of high-rate sludge are difficult (17). To obviate this difficulty, Sawyer (9) has suggested that secondary digesters in high-rate digestion systems be two to four times the capacity of high-rate primary digesters, to provide adequate time for solids separation and conditioning. This technique has been reported at Grand Rapids, Michigan (18), where the volatile solids loading to the primary digester is in excess of 0.25 pcf/day. The ratio between secondary and primary digester capacity is 3.5:1. In this case, secondary digester underflow solids exceed 10 percent and supernatant solids average less than 2 percent of the raw solids load. This indicates that the economics of decreasing the detention time in the primary digesters should be weighed against providing the additional capacity in the secondary digesters when solids dewatering is required.

11.2.5 Anaerobic Digester Supernatant

Anaerobic digester supernatant constitutes a high-strength recycle flow, as indicated by Table 11-3. It imposes a significant incremental organic loading on biological treatment processes. Separate supernatant treatment may be necessary in cases where available treatment capacity is limited.

TABLE 11-3
TYPICAL PROPERTIES OF ANAEROBIC
DIGESTER SUPERNATANT (19)

Parameter	Low Rate mg/l	High Rate mg/l
Total Solids	4,000 - 5,000	10,000 - 14,000
SS	2,000 - 3,000	4,000 - 6,000
BOD	2,000 - 3,500	6,000 - 9,000
VSS	650 - 1,000	2,400 - 3,000
Alkalinity (MO)	1,000 - 2,400	1,900 - 2,700
H_2S	70 - 90	190 - 440
NH_4 - N	240 - 560	560 - 630
pH	7.0 - 7.6	6.4 - 7.3

11.2.6 Impact of Alum and Iron Phosphorus Sludges on Anaerobic Digestion

The addition of alum or ferric chloride for phosphorus removal during biological treatment substantially increases the weight of the solids generated by the treatment process (20) (21) (22). However, in most instances, due to a higher sludge density, the volume of secondary sludge after chemical addition does not increase in the same proportion. If the chemical is added to the primary clarifier, the primary sludge concentration may be decreased. Piping, pumping and process units should be sufficient in size and number to handle the change in weight and/or volumes of sludges that result from chemical addition to either primary or secondary treatment processes.

Many studies have indicated that neither ferric chlorine nor alum phosphorus sludges inhibit anaerobic digestion (20) (21) (22) (23) (24). However, digester performance may be altered due to changes in physical sludge characteristics or feed sludge alkalinity. Studies have also indicated that chemically bound phosphorus is not released to the supernatant in the anaerobic digestion process (20) (24) (25).

At Chapel Hill, North Carolina (26), it was reported that digester alkalinity was reduced in the primary digester from 2,500 mg/l to 1,500 mg/l as $CaCO_3$ after addition of alum to one of two parallel trains (see Chapter 4), resulting in a need to add lime on one occasion. Further, the secondary digester underflow concentration decreased from a normal range of 6 to 7 percent to 3.0 percent with a coincident increase in the supernatant SS concentration from 1,000 mg/l to 10,000 mg/l. In spite of these difficulties, the digestion process itself produced a normal reduction in VSS throughout the entire alum treatment study.

11.2.7 Process Design and Cost Estimates

The following example illustrates the upgrading of a two-stage low rate digestion facility to a two-stage high rate digestion system. This was required due to the increase in plant flow from 1 to 3 mgd. A gravity thickener was added prior to digestion. Primary digester performance was improved by adding gas mixing and installing external heat exchangers to control the temperature more accurately. The comparison between existing and upgraded design conditions is presented in Table 11-4. The flow diagram of the upgraded plant is shown on Figure 11-4.

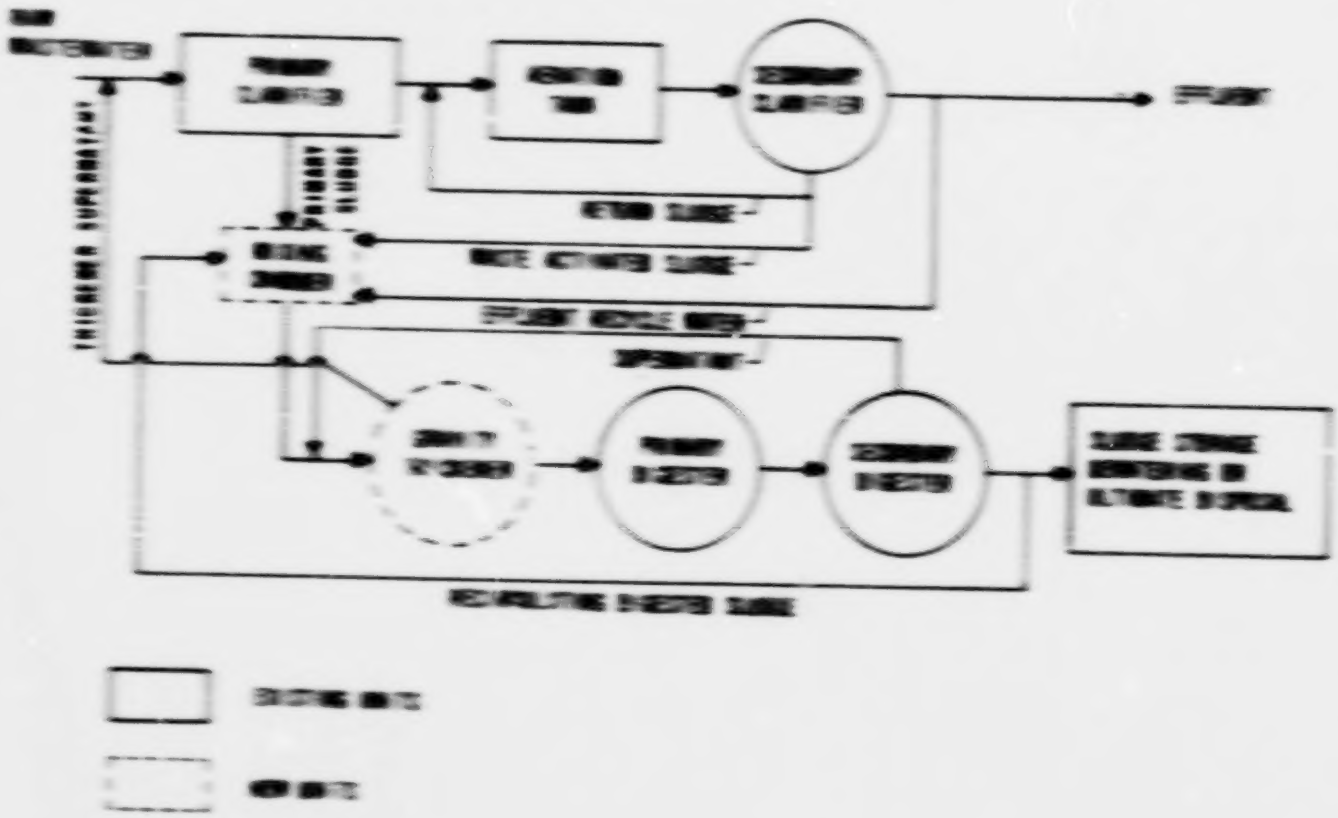
A mixing chamber ensures proper blending of sludges and effluent water prior to thickening. With thickening, the solids concentration and the volatile solids loading to the primary digester were increased from 2 to 5 percent and from 0.036 to 0.100 pcf/day, respectively. Provision was made for recirculating digested sludge back to the thickener at a rate of 50 percent of the raw sludge feed.

TABLE II-4
UPGRADING AN EXISTING LOW-RATE DIGESTION SYSTEM USING PRETHICKENING OF THE COMBINED SLUDGE AND IMPROVEMENTS TO THE PRIMARY DIGESTER

Parameter	Original Design	Upgraded Design
Plant Flow, mgd	1	3
Combined Sludge Characteristics		
Volume, gpd	8,900 (2%)	27,300 (2%)
Solids Contribution, lb/day	1,500	4,550
VSS, percent	70	70
Gravity Thickener		
Number	--	1
Solids Loading, lb/sf/day	--	10
Effluent Water Required to Dilute Sludge, gpd	--	250,000
Hydraulic Loading, gpd/sq ft	--	600
Thickened Sludge Volume, gpd	--	10,920 (5%)
Digester		
Number - Primary Digesters	1	1
Number - Secondary Digesters	1	1
Primary Digester Characteristics	Limited heating and mixing	New gas mixing and improved heating
Secondary Digester Characteristics	No heating or mixing	No heating or mixing
Digester Volume (Total), cu ft	60,000	60,000
Hydraulic Detention Time, days (Total)	50.4	41.0
days (Primary)	25.2	20.5
VSS Loading, lb/cf/day (Primary)	0.035	0.106

FIGURE II-4

UPGRADING AN EXISTING LOW-RATE DIGESTION SYSTEM USING PRE-THICKENING OF THE COMBINED SLUDGE AND IMPROVEMENTS TO THE PRIMARY DIGESTER


RAW WASTEWATER
PRIMARY CLARIFIER
AERATION TANK
SECONDARY CLARIFIER
EFFLUENT
PRIMARY SLUDGE
RETURN SLUDGE
MIXING CHAMBER
WASTE ACTIVATED SLUDGE
EFFLUENT RECYCLE WATER
SUPERNATANT
THICKENER SUPERNATANT
GRAVITY THICKENER
PRIMARY DIGESTER
SECONDARY DIGESTER
SLUDGE STORAGE DEWATERING OR ULTIMATE DISPOSAL
RECIRCULATING DIGESTED SLUDGE
EXISTING UNITS
NEW UNITS

The capital cost of this upgrading was estimated at $137,000, allocated as follows:

Thickener	$ 74,000
Digester Renovation	63,000
Total	$137,000

11.3 Aerobic Digestion

Aerobic digestion can be applied to the stabilization of primary sludges or combinations of primary and secondary sludges. It has been indicated that aerobic digestion is competitive with anaerobic digestion for activated sludge plants up to a size of at least 8 mgd (37). Fewer operational problems are associated with aerobic digestion than with anaerobic digestion. Hence, less laboratory control and daily maintenance are required. Also, the dangers of gas explosions are eliminated because the only gaseous byproducts of aerobic digestion are carbon dioxide and water vapor.

Typical concentrations of various constituents present in aerobic digester supernatant liquors are listed in Table 11-5. Compared to anaerobic digester supernatant, the recycle stream from aerobic digesters is relatively innocuous. The recycled SS, although high in concentration, consists of stabilized solids which can be effectively removed in primary settling and would have little impact on biological processes. The true organic loading, attributable to this recycle, is represented by the soluble BOD and is similar in strength to primary effluent.

TABLE 11-5
TYPICAL PROPERTIES OF AEROBIC DIGESTER SUPERNATANT

Parameter	Average mg/l	Range mg/l	Reference
SS	3,400	46 - 11,500	38
BOD	500	9 - 1,700	38
Soluble BOD	51	4 - 183	34
COD	2,600	288 - 8,140	38
Alkalinity	-	473 - 514	39
Kjeldahl N	170	10 - 400	38
Total P	98	19 - 241	38
Soluble P	26	2.5 - 64	38
pH	7.0	5.9 - 7.7	38

11.3.1 Process Considerations

Aerobic digestion is accomplished by aerating sludge until it is stable and relatively nuisance free. In the aerobic digestion of sludges, two different forms of oxidation take place, as shown in the following two reactions. First, a portion of the organic substrate in the sludge is oxidized and the remainder is converted to cellular matter. Second, the cellular matter produced is oxidized until only a relatively inert fraction remains.

$$(1)\quad \text{organic matter} + O_2 \longrightarrow \text{cellular matter} + CO_2 + H_2O$$

$$(2)\quad \text{cellular matter} + O_2 \longrightarrow \text{digested sludge} + CO_2 + H_2O$$

The first reaction, biological oxidation, is most significant in the stabilization of sludges with a high fraction of primary sludge. The second reaction, endogenous respiration, predominates in the aerobic stabilization of waste activated sludge or trickling filter humus. Since biological oxidation requires a higher oxygen transfer rate than endogenous respiration, more air dissolution capacity must be provided for the stabilization of sludges containing significant quantities of primary sludge.

Both pH and alkalinity are reduced in a properly functioning aerobic digester when nitrification takes place. Nitrification lowers pH according to the following reaction:

$$NH_4^+ + 1.5\ O_2 \xrightarrow{Nitrosomonas} NO_2^- + 2H^+ + H_2O$$

The second step of nitrification is as follows:

$$NO_2^- + 0.5\ O_2 \xrightarrow{Nitrobacter} NO_3^-$$

Theoretically, 7.1 lb of $CaCO_3$ alkalinity are destroyed per pound of ammonia nitrogen oxidized, since the two protons released neutralize one mole of $CaCO_3$ according to the following reaction:

$$2H^+ + CaCO_3 \longrightarrow H_2CO_3 + Ca^{2+}$$

In batch-type aerobic digestion, it is possible that the pH may drop to a rather low value (5.5) at increased detention times. However, this does not seem to inhibit the process (40).

Batch-type aerobic digestion operating data shown in Table 11-6 indicate process performance and the relationship between ammonia, nitrite, and nitrate nitrogen as a function of detention time. Table 11-6 also indicates that in the normal temperature range of 15 to 35 deg C, an increase in temperature increases the rate of VSS reduction.

TABLE 11-6
BATCH-TYPE AEROBIC SLUDGE DIGESTION OPERATING DATA FOR MIXTURES OF PRIMARY AND WASTE ACTIVATED SLUDGE (31)

Detention Time days	Temperature deg C	VSS Reduction percent	pH	Alkalinity mg/l	NH_3-N mg/l	NO_2-N mg/l	NO_3-N mg/l
5	15	21	7.6	510	54	Trace	None
10	15	32	7.6	300	3.2	1.20	64
30	15	40.5	6.6	81	4.0	0.06	170
60	15	46	4.6	23	20	0.30	335
5	20	24	7.6	590	54	Trace	None
10	20	41	7.6	390	4.9	0.59	60
15	20	43	7.8	560	7.0	2.27	29
30	20	44	5.4	31	20	0.19	275
60	20	46	5.1	35	7.0	0.51	700
5	35	36	7.9	630	14	0.10	None
10	35	45	8.0	540	10.0	0.08	None

Pure oxygen rather than air may be used in the aerobic digestion process. Because air contains only about 20 percent oxygen, the oxygen solubility in a system using air is only 1/5 that obtained when using pure oxygen. Consequently, pure oxygen may be used to stabilize thicker sludges in which the high oxygen uptake rates cannot be satisfied with air aeration.

A three month plant scale study of aerobic digestion using pure oxygen in a covered system was conducted at Speedway, Indiana (32). Aeration was accomplished in a covered 11,000 cu ft, four-stage reactor followed by a clarifier for decanting the sludge. The study was divided into two phases, the first treating only waste activated sludge and the second treating mixed primary and waste activated sludge. Much of the heat generated by biological oxidation was retained within the covered system. This resulted in a significant increase in sludge temperature and a corresponding increase in the rate of VSS destruction. The results of this study are shown in Table 11-7. A positive DO level of 2 mg/l was required for the process to function well.

TABLE 11-7
RESULTS OF HIGH PURITY OXYGEN AEROBIC DIGESTION
SPEEDWAY, INDIANA (32)

Parameter	Phase 1	Phase 2
Feed Sludge Type	WAS[1]	57% WAS + 43% PRI.[2]
Feed TSS, percent	2.14	3.06
VSS/TSS	0.77	0.66
Biodegradable VSS/TSS	0.45	0.40
Feed Temperature, deg C	19.5	16.0
Average Ambient Air Temperature, deg C	7.5	-2.2
Sludge Temperature - Stage 4, deg C	33.0	31.0
HRT, days	16.3	11.6
Volatile Solids Loading Rate, pcf/day	0.064	0.109
VSS Reduction, percent	44	43
Biodegradable VSS Reduction, percent	94.6	89.6

[1] Waste Activated Sludge
[2] Primary Sludge

Batch tests were conducted at the Metropolitan Denver Plant on a pure oxygen aerobic digestion system using open tanks and a special oxygen diffuser system producing extremely fine bubbles (33). These tests indicated that a 40 to 60 percent reduction in VSS can be obtained with an HRT of ten days, when digesting waste activated sludge.

The overall cost effectiveness of using high purity oxygen for aerobic digestion is greatly enhanced if the wastewater treatment plant employs the high purity oxygen activated sludge process.

11.3.2 Design Basis

There are two methods of operating aerobic digestion tanks. One is on a continuous basis, and the other is on an intermittent batch basis. The latter method is used most frequently in smaller plants. Continuous digester operation requires provision for continuously decanting the supernatant, but offers the advantages of equalizing air requirements and providing a slow continuous supernatant return. In the batch operation, the sludge is supplied with air in a completely mixed tank for a protracted period, followed by quiescent settling and decantation as new sludge is admitted. However, care must be taken in cold climates to avoid creating a slush at prolonged detentions and subfreezing air temperatures.

In designing an aerobic digestion system, the characteristics of the raw sludge must be fully identified. This can best be done with pilot plant studies. A small (55-gallon) aerobic digestion pilot plant can be used for this purpose. Through such studies, the detention time, oxygen requirements and optimum VSS reduction can be determined. Mathematical models analyzing aerobic digestion kinetics have been reported by Reynolds (34) and by Smith, et al (35).

Table 11-8 contains a summary of parameters used in the design of aerobic digestion units for municipal wastewater sludges. When considering aeration requirements, recognition must be given to the oxygen demand of the ammonia nitrogen present.

11.3.4 Use of Aerobic Digestion for Upgrading Sludge Handling Facilities

11.3.4.1 Use of Existing Facilities

As an upgrading technique, aerobic digestion can be carried out in existing unused tankage, such as old Imhoff tanks or old clarifiers. If unusually shaped basins are used, attention should be placed on ensuring that complete mixing will be achieved and that dead spots will be eliminated. Potential dead spots can be filled and covered with concrete. Air diffusion systems are more easily adapted to unusual basin shapes than are surface aerators.

11.3.4.2 Supplemental Aerobic Digestion

Aerobic digestion can be used in conjunction with existing anaerobic digesters. In Monroe, Wisconsin, and Corpus Christi, Texas, it has been found best to digest the primary sludge anaerobically and the waste biological sludge aerobically (36). The advantage of this segregation is that the primary sludge is not diluted by the waste biological sludge and the anaerobically digested primary sludge subsequently filters better on a vacuum filter when it does not contain the waste biological sludge.

11.3.4.3 Conversion of Anaerobic Digesters to Aerobic Digesters

If existing anaerobic digesters are overloaded and for some reason cannot be upgraded as described in Section 11.2, they can be converted to aerobic digesters. Aerobic digesters yield similar VSS reductions and are relatively odor free. Hence, this conversion may be applicable to small overloaded plants in residential areas. Aerobic digestion will usually require an increase in the blower capacity of the air supply system, which would significantly increase the yearly operating cost. However, these increased costs could be offset by savings in maintenance requirements. An alternative method of supplying the additional air is by using mechanical surface aerators where tank geometry permits.

TABLE 11-8
AEROBIC DIGESTION DESIGN PARAMETERS USING AIR

Parameter	Value	Remarks	Reference
Solids Retention Time	10-15[1] 15-20[2]	Depending on temperature, type of sludge, etc.	[illegible] [illegible]
Volume Allowance, cu ft/capita	3-4		[illegible]
VSS Loading, pcf/day	0.024-0.14	Depending on temperature, type of sludge, etc.	[illegible]
Air Requirements			
Diffuser System, cfm/1,000 cu ft	20-35[1]	Enough to keep the solids in suspension and maintain a DO between 1-2 mg/l	[illegible]
cfm/1,000 cu ft	≥60[2]		[illegible]
Mechanical System, hp/1,000 cu ft	1.00-1.25	This level is governed by mixing requirements. Most mechanical aerators in aerobic digesters require bottom mixers for solids concentration greater than 8,000 mg/l, especially if deep tanks (>12 feet) are used	
Minimum DO, mg/l	1.0-2.0		[illegible]
Temperature, deg C	>15	If sludge temperatures are lower than 15 deg C, additional detention time should be provided so that digestion will occur at the lower biological reaction rates	
VSS Reduction, percent	[illegible]		[illegible]
Tank Design		Aerobic digestion tanks are open and generally require no special heat transfer equipment or insulation. For small treatment systems (0.1 mgd), the tank design should be flexible enough so that the digester tank can also act as a sludge thickening unit. If thickening is to be utilized in the aeration tank, sock type diffusers should be used to minimize clogging	[illegible]
Power Requirement, [illegible] Population Equivalent	[illegible]		[illegible]
Power Cost			
$/yr/lb BOD Removed	[illegible]	These cost data are based upon three operational plants in Pennsylvania and do not reflect current escalation of energy costs	[illegible]
$/yr/capita	[illegible]		

[1] Waste activated sludge alone

[2] Primary and waste activated sludge or primary sludge alone

When existing anaerobic digesters are converted to aerobic digesters, the impact of retaining or removing the digester covers on sludge temperature and the temperature of the air under the covers, accessibility of diffused air equipment, generation of objectionable odors during warm weather periods, oxygen demand and air supply capacity must be considered. Experience at one midwestern city indicated that the combination of an air diffusion system and a covered aerobic digester (converted from an existing anaerobic digester) increased the air temperature inside the digester to 115 deg F and the sludge temperature to 97 deg F (15). At these high temperatures, objectionable odors were produced due to high concentrations of organic sulfides in the sludge and the low digester DO. Under these conditions the sulfur-bearing compounds were not completely oxidized resulting in the release of H_2S to the atmosphere. The increased oxygen demand of the sludge that occurred at the unanticipated high temperatures also reduced the ability of the air supply system to maintain the desired DO levels.

11.3.4 Process Design and Cost Estimate

The following example illustrates upgrading sludge handling facilities with aerobic digestion. Due to continued operational problems with an existing anaerobic digestion facility, a community decided to convert the existing anaerobic digesters to aerobic digesters by the installation of blower and diffused air equipment. The waste activated sludge from the 4 mgd activated sludge plant is settled along with raw wastewater in the primary clarifiers and pumped directly to the digestion system at an average solids concentration of 4 percent. The upgraded facility is shown on Figure 11-5.

In the upgraded system, the two aerobic digesters will be operated continuously at a total detention time of 17.4 days. This detention time is in excess of the 15 to 20 days required and will allow the aerobic digestion process to handle increased solids loading in the future. A summary of the design data for the upgraded facility is presented in Table 11-9.

The estimated total capital costs for the upgrading are $203,000, broken down as follows:

Air System	$157,000
Renovation to Existing Tank	46,000
Total	$203,000

11.4 Heat Treatment of Sludge

The heat treatment process involves heating sludge for short periods of time under pressure. It is essentially a conditioning process which prepares sludge for dewatering without the use of chemicals. In addition, the sludge is sterilized and generally rendered inoffensive.

FIGURE 11-5

CONVERSION OF ANAEROBIC TO AEROBIC DIGESTION

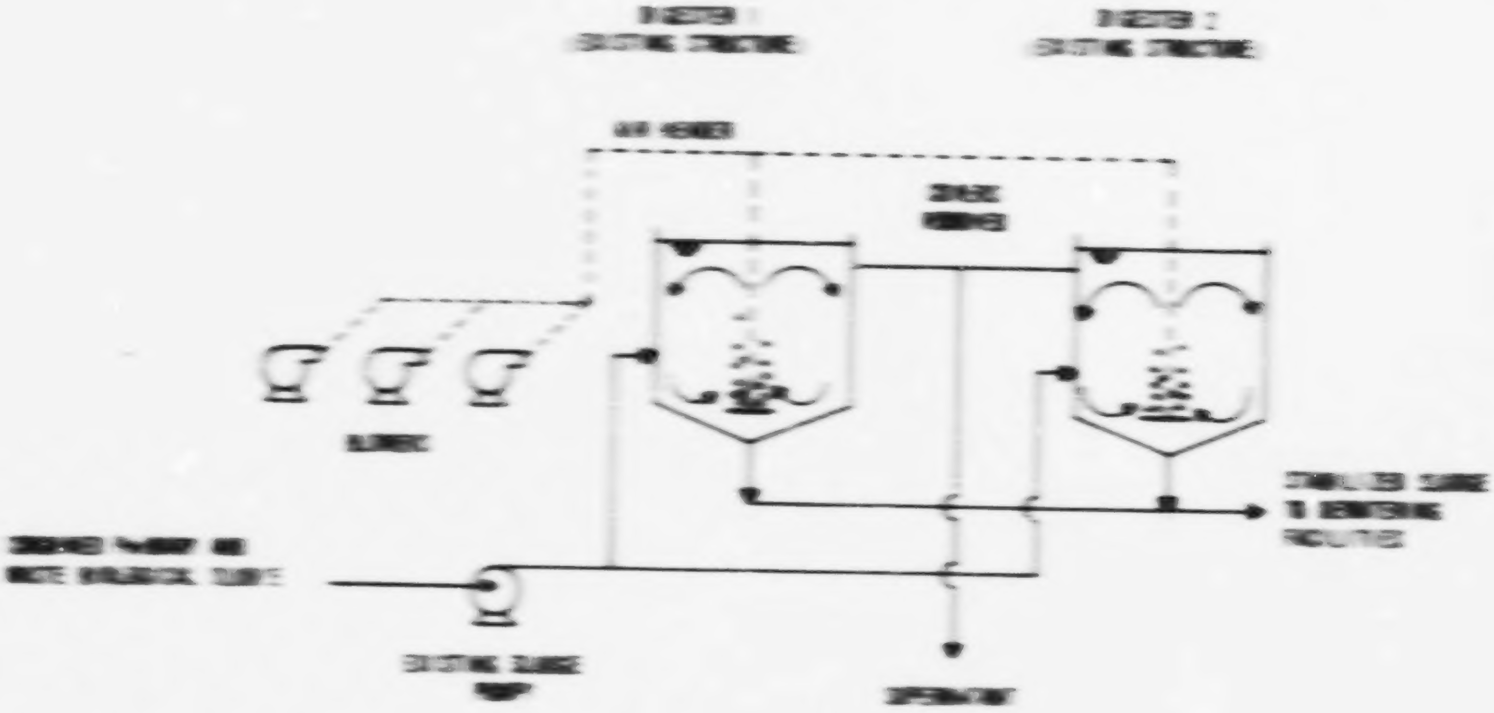

DIGESTER 1
(EXISTING STRUCTURE)
DIGESTER 2
(EXISTING STRUCTURE)
AIR HEADER
BLOWERS
STABILIZED SLUDGE
TO DEWATERING
FACILITIES
THICKENED PRIMARY AND
WASTE BIOLOGICAL SLUDGE
EXISTING SLUDGE
PUMP
SUPERNATANT

TABLE 11-9
AEROBIC DIGESTER UPGRADING DESIGN PARAMETERS

Description	Value
Plant Design Flow, mgd	4
Number of Digesters	2
Volume (each), cu ft	75,000
Total Solids Added to Aerobic Digesters, lb/day	7,500
Volatile Solids Added to Aerobic Digesters (70 percent volatile), lb/day	5,250
Sludge Volume (3 percent solids), gpd	30,000
VSS Loading (each) pcf/day	0.035
Hydraulic Detention Time (total), days	37.4
Aeration Requirements (each), cfm	4,500[1]

[1] (60 cfm/1,000 cu ft)

When sludge is conditioned by heat treatment, thermal activity releases the bound water and results in coagulation of solids. In addition, hydrolysis of proteinaceous material occurs resulting in cell destruction and ammonia nitrogen release. Approximately 30 to 40 percent of the VSS are solubilized resulting in a very high strength recycle liquor.

Two variations exist for heat treatment. In both systems, sludge is passed through a heat exchanger into a reactor vessel, where steam is injected directly into the sludge to bring the temperature to within the range of 140 to 200 deg C, under a pressure of 150 to 300 psig. In one variation air is also injected into the reactor vessel with the sludge. The detention time in the reactor is approximately 30 minutes. After heat treatment, the sludge passes back through the heat exchanger to recover heat, and then is discharged to a thickener-decant tank. The thickened sludge may be dewatered by vacuum filtration or centrifugation to a solids content of 40 to 45 percent.

The supernatant and filtrate recycle liquor contains high concentrations of short chain water soluble organic compounds and ammonia nitrogen. The strength of the recycle is much higher than that from anaerobic or aerobic digestion, but is generally amenable to biological treatment. This flow should be returned to the plant for treatment but careful consideration must be given to the effects of the increased organic and ammonia nitrogen loads. Otherwise, the liquor must be treated separately and then returned to the plant. A complete discussion of the heat treatment process is presented in the Process Design Manual for

Sludge Treatment and Disposal (1). Characteristics of filtrate for a heat treated sludge with and without air addition are listed in Table 11-10. The results presented are from a pilot study on high-purity oxygen activated sludge with a feed concentration of 2.68 percent from the Batavia, New York, Wastewater Treatment Plant.

TABLE 11-10
CHARACTERISTICS OF HEAT TREATED SLUDGE FILTRATE AT BATAVIA (46)

Parameters	With Air Injection mg/l	Without Air Injection mg/l
COD	17,000	15,900
BOD	7,210	7,650
TSS	8,000	11,400
VSS	7,000	10,200
Total P	200	140
Total N	1,610	1,510
NH_3 - N	740	540
pH	4.7	6.4

The advantages claimed for the heat treatment process are:

1. Excellent dewatering characteristics of treated sludge without chemical conditioning.
2. Innocuous sludge suitable for ultimate disposal by a variety of methods.
3. Few nuisance problems.
4. Suitable for many types of sludge which cannot be stabilized biologically.
5. Reduction in subsequent incineration energy requirements.
6. Reduction in size of subsequent vacuum filters and incinerators.

Disadvantages claimed for the heat treatment process are:

1. High construction and operation
2. Specialized supervision and maintenance requirements due to high temperatures and pressures
3. High recycle organic and ammonia nitrogen loads
4. Expensive material costs to prevent corrosion.

11.5 Lime Stabilization of Sludge

Land disposal of raw sludge is objectionable primarily because the sludge contains a large quantity of pathogenic microorganisms and heavy metals. Furthermore, biological conversion of the sludge cake into an inert material can create odor problems. Lime stabilization of raw sludge can improve its acceptability for land disposal.

Various investigators have reported that the addition of lime to raw or digested sludges to pH ranges of 10.2 to 12.5 has effectively reduced the number of pathogenic organisms present (47) (48) (49). Current U. S. EPA sponsored work indicates that the pH should be increased to 12.0 for more effective disinfection (1). Farrell, et al. have investigated the bactericidal effect of lime on chemically precipitated primary sludges. Their findings are presented in Table 11-11.

TABLE 11-11
BACTERICIDAL EFFECT OF LIME ADDITION TO
CHEMICALLY PRECIPITATED SLUDGES (50)

Sludge	Bacterial Count (organisms/liter of sludge)		
	Salmonella Species	Pseudomonas aeruginosa	Total Aerobic Count
Alum Primary	110	1,300	41×10^6
Limed Alum Primary	None detected	None detected	5.0×10^6
Ferric Primary	> 24,000	610	190×10^6
Limed Ferric Primary	None detected	None detected	0.29×10^6

Lime addition to raw sludge does not render it permanently stable because the pH eventually drops and surviving organisms, or organisms that recontaminate the sludge, can create nuisance conditions (30). Disposal in sanitary landfills of dewatered lime stabilized raw sludge has been suggested as a satisfactory procedure. In such cases, the sludge should be deposited in thin layers and covered daily.

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CHAPTER 12

SLUDGE DEWATERING

12.1 General

The selection of a sludge dewatering process is mainly dependent on the subsequent treatment and disposal of the dewatered sludge. If incineration is used, sludges with low moisture content are required to minimize fuel consumption. However, if dewatered sludge is taken directly to a landfill, higher sludge moisture contents may be tolerable depending on the construction and management practices of the landfill operation. Other factors that should be considered are space limitations, economics and general operation.

12.2 Vacuum Filtration

Vacuum filtration is the most common mechanical method of sludge dewatering utilized in the United States. The majority of vacuum filters are installed in larger facilities where sludge drying beds are impractical due to limited space, or where incineration is necessary for maximum volume reduction. In small plants, vacuum filters have not been commonly used because of the higher operating skill required; the increased construction costs for facilities to house chemical storage, conditioning and filtration facilities; and the higher operating cost for chemical conditioning. Today, however, there is an increasing trend toward the use of vacuum filters in smaller communities, because of improvements in operator training, the increasing scarcity of land and, in some instances, the difficulty of obtaining unskilled labor for cleaning of sludge drying beds. A flow diagram of a typical vacuum filter installation is shown on Figure 12-1.

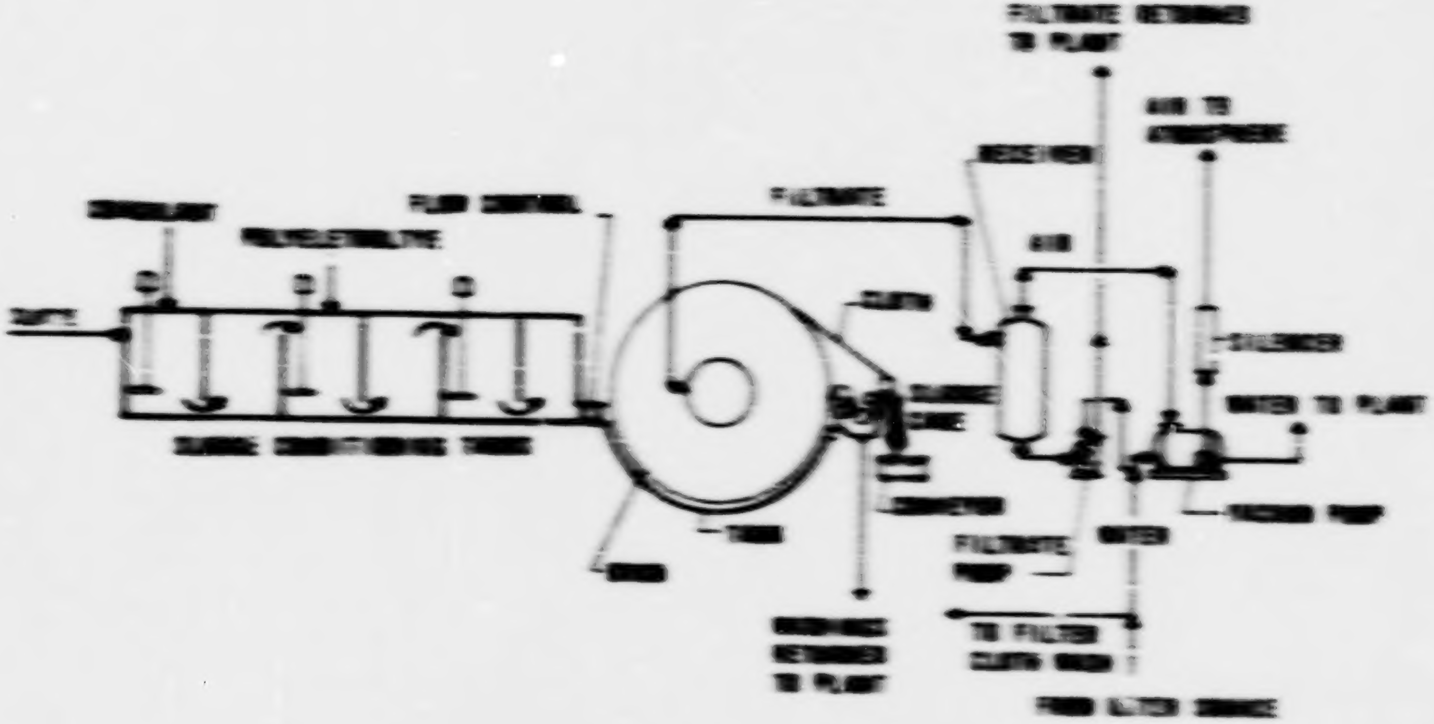
12.2.1 Process Considerations

Process operating considerations for vacuum filtration include:

1. Type of sludge
2. Feed solids concentration
3. Sludge age
4. Chemical requirements for sludge conditioning
5. Sludge mixing and flocculation
6. Drum speed and drum submergence
7. Filter fabric characteristics
8. Filter cake discharge.

FIGURE 12-1

TYPICAL VACUUM FILTER FLOW DIAGRAM



FILTRATE RETURNED TO PLANT
AIR TO ATMOSPHERE
RECEIVER
FILTRATE
AIR
SILENCER
COAGULANT
POLYELECTROLYTE
FLOW CONTROL
SLUDGE
CLOTH
SLUDGE CAKE
WATER TO PLANT
SLUDGE CONDITIONING TANKS
CONVEYOR
VACUUM PUMP
TANK
DRUM
FILTRATE PUMP
WATER
DRAINAGE RETURNED TO PLANT
TO FILTER CLOTH WASH
FROM OUTSIDE SOURCE

Each of these parameters affects the filter yield, economy of operation and filter cake characteristics. The nature of the sludge to be filtered is important in establishing several of the design parameters. Sludge particle size, shape, consistency and density affect filterability and chemical conditioning requirements (1). Because digestion results in the breakdown of coarse solids into finer particles, vacuum filtration of raw primary or primary-secondary sludges permits higher filter yields, lower chemical requirements and lower cake moisture contents than vacuum filtration of digested sludges. As the ratio of secondary to primary sludge increases, vacuum filtration of undigested sludge becomes more difficult.

The feed solids concentration has an important influence on the filtration rate and filter yield. In a large number of cases where poor filter performance has been experienced, sufficient care in operation has not been given to optimizing the feed solids concentration (1). The optimum solids concentration for filtration is 8 to 10 percent. Higher solids content makes the sludge difficult to pump and condition for dewatering. Lower solids content requires the use of more or larger vacuum filters than necessary. In general, sludge yield increases directly in proportion to an increase in feed solids concentration, although data have been reported where the increase was more than proportional (2)(3)(4).

It is generally agreed that as raw sludge ages, it becomes more difficult to dewater on vacuum filters. This effect has been reported by Ettelt and Kennedy (5), and Tenney, et al. (3) who observed increasing filter cake moisture with increasing age of thickened activated sludge. Tests at the Chicago Sanitary District (5) showed that even a short holding period after thickening can significantly affect vacuum filter performance. It was observed that the filter cake solids concentration varied linearly with detention time, decreasing from 17 percent for immediate filtering after thickening to 13 percent after a holding period of 3 1/2 hours. In this situation, it was found that freshening of the sludge by reaeration prior to filtration dramatically reduced the filter cake moisture and the required ferric chloride conditioning dosage.

Few, if any, raw or digested wastewater sludges can be successfully dewatered without some form of chemical conditioning using ferric chloride, lime and/or polyelectrolytes. Proper sludge conditioning requires a determination of optimum chemical dosages. Experience and careful laboratory monitoring of the sludge characteristics are key factors in maintaining the proper chemical proportions and concentrations. Where a mixture of undigested primary and secondary sludge is to be dewatered, it is essential that careful attention be given to maintaining a uniform ratio of primary to secondary sludge. Mechanically mixed and/or aerated blending tanks are often provided for this purpose.

Optimum sludge mixing and flocculation under varying conditions require that sludge conditioning tanks be provided with variable-speed mixer drives, adjustable weirs or flow control devices to vary the sludge detention time and multiple points of chemical application as indicated on Figure 12-1. The sludge slurry must be agitated sufficiently to maintain uniformity; however, excessive agitation should be avoided so that the conditioned slurry particles are not sheared.

Drum speed and drum submergence are important factors in the operation of vacuum filters since they affect filter yield and filter cake moisture. Increasing the drum submergence generally results in increased filter yield, but produces a filter cake with higher moisture content. Decreasing the drum speed, i.e., increasing the cycle time, has the opposite effect of decreasing the filter yield, but produces a cake with lower moisture content.

Much information is available from the various manufacturers of vacuum filters and filter media on the selection of proper media. A wide variety of media are now available including natural and synthetic fabrics and metal coils. Because of the variable nature of sludge, it is important that the selection of the media be based as nearly as possible on actual plant conditions. Often this can best be accomplished through laboratory experimentation, using the Filter Leaf Test, to provide information on filter fabric blinding, filtrate quality and cake characteristics.

In general, a media is selected on the basis of the cleanest possible filtrate consistent with high filtration rates, and where incineration is used, lowest moisture contents. In incineration situations, it is often economically justifiable to oversize the filters to achieve lower moisture contents and significant fuel savings. The concentration of total solids in the vacuum filter filtrate is the primary measure of filtrate quality. The SS may vary between 100 and 20,000 mg/l depending on sludge type, the filter media, and the applied vacuum (1). In practice, filtrate SS concentrations of 500 to 2,000 mg/l are normal. Although the solids in the filtrate normally resettle readily when returned to the head of the plant, fine solids may resist removal and be continuously recycled. Accumulation of these solids in the system may eventually reduce overall plant efficiency. Activated sludges normally contain more fine particles than primary sludge and, therefore, require tighter filter medias and slower filtration rates.

A prerequisite for good filter operation is that the sludge cake discharge cleanly from the filter media (6). Characteristics of both the sludge and the media affect filter cake discharge. Primary sludges having high fiber content generally produce a drier cake that discharges more readily than activated or digested sludges containing finer material. Because the cake discharge characteristics of various fabrics differ, a number of materials should be evaluated when running Filter Leaf Tests to determine the best fabric for a given sludge. Generally, cake discharge occurs through the flexing action of the media passing around a small diameter discharge roller. Where sludge characteristics are such that the cake tends to adhere to the media, a scraper blade may be provided to assist removal.

12.2.2 Evaluation of the Vacuum Filtration Process

Experience has shown that there are considerable variations in filtration rate, not only between different sludge types but also between the same types of sludges at different times and locations. The discrepancies in filter test results at different plants are usually related to variations in feed solids concentrations, particle size distributions, and industrial waste components in the raw wastewater. Obviously then, it is of major concern to have a laboratory technique which can accurately predict the performance of a full-scale vacuum filter prior to its installation.

The Buchner Funnel Test and the Filter Leaf Test are commonly used in laboratory testing programs for determining the filterability of sludges. When the amount of representative sludge is limited (less than 10 liters), it is advisable first to perform the Buchner Funnel Test to determine optimal chemical dosage and sludge filtration characteristics. The Filter Leaf Test can then be run at the optimum condition to determine filter yield. If a large amount of sludge is available, the Buchner Funnel Test can be eliminated and the Filter Leaf Test run instead.

The main purpose of the Buchner Funnel Test is to determine optimum chemical requirements based on a comparison of the specific resistance of chemically treated sludge with that of untreated sludge. An approximate filter yield can also be calculated from the Buchner Funnel Test. Basically, the Buchner Funnel Test consists of filtering 100 ml of sludge, either raw or conditioned, through filter paper under a vacuum of 20 to 25 inches of mercury. The volume of filtrate (V, in ml) with time is noted and plotted against elapsed time/volume (t/V in sec/ml) to obtain the slope of the resulting line. Using the above information, the specific resistance of the sludge is calculated from the following equation:

$$r = 6.91 \times 10^{6} \frac{bA^{2}P}{\mu c}$$

where:

r = Specific resistance, sec^{2}/g
b = Slope of plot (V vs. t/V), sec/ml^{2}
A = Area of filter, sq cm
P = Filtration vacuum, psig
μ = Absolute viscosity of filtrate, centipoise
c = Initial SS concentration, mg/ml

The dimensions of the variables in the above equation are in units typically measured in the laboratory. The conversion constant, 6.91×10^{6} reduces the variables to units which are dimensionally consistent. The specific resistance as calculated by the above equation would be expressed as sec^{2}/g. Typical values of specific resistance for various sludges are shown in Table 12-1.

TABLE 13-1
TYPICAL SPECIFIC RESISTANCE VALUES FOR VARIOUS SLUDGES

	Primary Sludge Specific Resistance, 10^7 sec^2/g	Activated Sludge Specific Resistance, 10^7 sec^2/g	Reference
Raw Sludge	1,300-2,100	2,800	7
5% $FeCl_3$	10	140	8
5% $FeCl_3$ + 5% CaO	3	60	8
2% Polyelectrolyte	3	50	8
Digested Sludge	400-1,600	800	7
5% $FeCl_3$	25	–	8
5% $FeCl_3$ + 5% CaO	10	–	8
2% Polyelectrolyte	30	–	8

The approximate filter yield can be estimated from the specific resistance data using the following relationship (9):

$$L = 0.0357 \left(\frac{100 - C_i}{C_i - C_f}\right)\left(\frac{mP (100 - C_i)}{\theta R \mu}\right)^{1/2}$$

where:

L = Filter yield, psf/hr
C_i and C_f = Initial and final moisture content of the sludge, percent
m = Percentage of time for which vacuum acts during cycle
θ = Time for one drum revolution, minutes (usually between 1.5 and 5 minutes)
μ = Absolute viscosity of filtrate, centipoise
R = $r \times 10^{7}$ g/sec^2
r = Specific resistance, sec^2/g
P = Filtration vacuum, psig

The Filter Leaf Test techniques are simple, and the test can be run with minimum effort. The advantage of the Filter Leaf Test is that the filter yield is measured and not merely calculated using an empirical equation. With careful laboratory techniques, results will be closely indicative of full-scale vacuum filter operation, although plant scale tests are more conclusive and should be used whenever possible. A scale-up design factor of 0.9 is typically used for the Filter Leaf Test where plant scale tests are not practical.

The Filter Leaf Test is usually performed on a 0.1 sq ft filter leaf. The main objective of the test is to evaluate the effect of different fabrics, fabric forms and drying times on filter yield. Varying doses and types of chemicals should also be tested to establish chemical conditioning requirements. The basic steps in performing a Filter Leaf Test are outlined below and should be repeated a number of times to optimize the performance of each media tested.

1. Condition the sludge in a 2 to 5 gallon container.

2. Submerge the filter leaf in the sludge slurry and apply vacuum for a fixed form time.

3. Gently remove the filter leaf from the sludge slurry to allow the cake to dry for a fixed drying time.

4. Remove the cake from the filter leaf and measure the weight and moisture content of the cake. Note cake thickness and separability from the media.

5. Measure the filtrate SS and filtrate volume.

The filter yield is then calculated using the following equation:

$$L = \frac{\text{dry sludge weight (g)} \times \text{number of filtration cycles/hr}}{453.6 \times \text{area of test filter leaf (sq ft)}}$$

where:

L = filter yield in psf/hr

12.2.3 Vacuum Filter Performance

The results of studies by the Dow Chemical Company (10) on vacuum filter performance are shown in Table 12.2. Various types of municipal sludges were dewatered using polyelectrolytes and inorganic chemicals as conditioners for the same sludges. The data shown in Table 12.2 indicate that the yield obtained when using polyelectrolytes for sludge conditioning is generally higher than that obtained when using inorganic chemical conditioners.

Vacuum filtration of pure oxygen waste activated sludge has been studied at Batavia, New York, and in several other demonstration plants (11) (12). Data from these studies are shown in Table 12.3. The settling characteristics of pure oxygen waste activated sludge are such that, where primary settling is omitted, it can be thickened adequately in the clarifiers for direct vacuum filtration. In this instance, additional sludge storage capacity beyond that

TABLE 12-3
VACUUM FILTRATION PERFORMANCE USING INORGANIC CHEMICALS AND PURIFLOC C-31 ON MUNICIPAL SLUDGE (10)

Type of Sludge	Filter Media	Chemical Conditioning	Dosage lb/ton dry solids	Solids Concentration Initial percent	Solids Concentration Final percent	Filter Yield psf/hr
1a Raw primary	Coil	$FeCl_3$	162.4			
		Lime	166.0	7.57	30.4	6.91
b Raw primary	Coil	C-31	11.0	7.0	30.0	7.33
2a Raw primary	Coil	$Fe_2(SO_4)_3$	60			
		Lime	100	9.4	24.0	8.6
b Raw primary	Coil	C-31	8.4	9.6	30.0	11.3
3a Raw primary	Open synthetic	$FeCl_3$	80.0			
		Lime	300.0	13	40.0	5.0
b Raw primary	Open synthetic	C-31	18.0	12	40.0	7.0
4a Raw primary	Open synthetic	$FeCl_3$	70.0			
		Lime	190.0	11.2	40.0	1.1
b Raw primary	Open synthetic	C-31	20.0	10.7	41.5	4.3†
5a Digested primary	Long napped dacron	$Fe_2(SO_4)_3$	(86.00/4.1)	7.2	36.0	4.6
b Digested primary	Long napped dacron	C-31	(85.50/4.1)	7.2	33.5	11.0
6a Digested primary	14 x 14 Saran	$FeCl_3$	66.0			
		Lime	200.0	13.0	14.0	9.2
b Digested primary	14 x 14 Saran	C-31	17.0	13.0	12.0	25.0
7a Elutriated/digested primary	Napped polyester	$FeCl_3$	36.0	6.4	46.0	3.74
b Elutriated/digested primary	Napped polyester	C-31	10.2	7.7	42.9	12.66
8a Elutriated/digested primary	Napped polyester	$FeCl_3$	100.0	10.4	32.7	4.00
b Elutriated/digested primary	Napped polyester	C-31	8.0	10.4	30.6	3.91
9a Elutriated/digested primary	Napped polyester	$Fe_2(SO_4)_3$	100.0	10.9	14.0	2.0
b Elutriated/digested primary	Napped polyester	C-31	9.0	11.4	15.0	3.5
10a Elutriated/digested primary and secondary	Long napped dacron	Lime and $Fe_2(SO_4)_3$	(80.90/7.5)	9.4	25.0	4.7
b Elutriated/digested primary and secondary	Long napped dacron	C-31	(80.72/4.1)	9.4	24.0	7.4
11a Digested primary and secondary	Synthetic	$FeCl_3$	1,400.0 (total)	4.4	30.2	3.2
		Lime	or ($10.07/ton)			
b Digested primary and secondary	Synthetic	C-31	10.0 or	4.3	33.7	3.6
			($6.12/ton)			
12a Elutriated/digested primary and secondary	Napped dacron	$Fe_2(SO_4)_3$	160.0			
		Lime	120.0	6.0	20.0	3.4
b Elutriated/digested primary and secondary	Napped dacron	C-31	22.0	9.0	25.0	7.25

† Yield intentionally kept down to avoid overloading incinerator.

TABLE 12-3
SUMMARY DATA ON VACUUM FILTRATION OF PURE OXYGEN AERATION SLUDGES (11)

Location	Type of Sludge	Feed Solids	Chemical Dose Rate % of Dry Solids FeCl$_3$	Chemical Dose Rate % of Dry Solids Lime	Cake Yield	Cake Solids
		percent			psf/hr	percent
Batavia, N. Y.[2]	Waste activated	2.75	17.0	22	1.96	20.1
	Waste activated	2.77	17.4	–	2.14	20.5
	Waste activated	2.29	19.2	–	4.35	16.0
	Thickened waste activated	4.27	20.0	–	3.42	14.5
Louisville, Ky.[3]	Waste activated	2.90	16.5	–	2.26	13.3
	Waste activated	3.20	19.0	–	1.95	17.6
	Combined[1]	6.30	8.0	25.0	4.00	29.0
New Orleans, La.[3]	Waste activated	2.00	10.0	–	3.60	16.3
	Waste activated	3.00	15.0	–	3.20	20.5
Philadelphia, Pa.[3]	Waste activated	1.40	14.6	–	1.72	18.8
	Waste activated	1.84	14.0	–	1.26	16.0

Notes:

1. Combined Sludge - 3 parts raw primary sludge, 1 part waste activated sludge.
2. Pilot Scale Filter (3 ft diameter by 1 ft length).
3. Filter Leaf (filter area = 0.1 sq ft).

available in the clarifiers is necessary to allow optimum filter operation without adversely affecting clarifier performance. The most effective chemical conditioner for vacuum filtering pure oxygen sludge at Batavia was found to be ferric chloride at a dosage of 175 to 200 lb/ton of dry solids (11). Cake yield was roughly doubled by increasing the drum speed from 6.3 minutes per revolution to 3.5 minutes per revolution, but cake moisture content increased substantially. The use of polyelectrolytes alone or in combination with inorganic chemicals was not as effective as ferric chloride.

12.3.4 Upgrading Existing Vacuum Filters

In some instances, recycling a load of fine solids within the plant can seriously impair vacuum filter performance. Such conditions are created by poor quality digester supernatant or elutriate which must be corrected before good filter performance can be reestablished.

The need to upgrade existing vacuum filters is usually due to an increase in sludge production or a change in sludge characteristics. In these cases, existing vacuum filter operations may be upgraded by improving one of the following:

1. Feed solids concentration
2. Filter media
3. Sludge conditioning

Sludge thickening to increase feed solids concentration to the vacuum filter is generally the most economical and therefore the most desirable method of increasing the filtration rate. An increase in feed solids concentration will normally result in at least a proportional increase in the filtration rate (1) (3). Therefore, the effect of upgrading sludge thickening facilities to increase feed solids concentration from 4 percent to 5 percent would be to increase the filtration rate or sludge dewatering capacity by 25 percent. For this reason, steps should be taken to ensure that feed solids concentration is maximized before other upgrading techniques are considered.

The use of proper vacuum filter media is very important to efficient filter performance. Most primary sludges and certain industrial waste sludges have fibrous, non-uniform solids that may clog or blind improper media and lead to lower filter yields, increased chemical consumption and the need for frequent washing (1). For this reason, primary sludges are most effectively dewatered with media having comparatively large openings that resist blinding. These media, however, are not suitable for activated or digested sludge due to the higher proportions of fine particles. These sludges require tighter media and lower filtration rates. Therefore, where sludge characteristics have been altered significantly due to upgrading procedures, vacuum filter operation often may be correspondingly upgraded by replacing filter media as indicated by laboratory Filter Leaf Tests or full-scale trials (1).

Chemical conditioning is normally required prior to vacuum filtration of sludge. The operation of a dewatering facility, therefore, may be upgraded by providing more efficient chemical sludge conditioning. Most older filter installations were designed to use inorganic chemicals, such as ferric chloride and lime, as sludge conditioners. Within the last 10 years, organic polyelectrolytes have begun to replace inorganic chemicals as sludge conditioners, and offer an attractive advantage in more economical storage, handling and feeding equipment. Polyelectrolytes are also less corrosive and frequently less expensive than inorganic chemicals. The use of polyelectrolytes in improving the operation of vacuum filters has been practiced at treatment plants in Bay City, Michigan (13), and Kansas City, Missouri (14), among others.

The Bay City Wastewater Treatment Plant provides primary treatment for 7.0 mgd and produces approximately 430 tons of dry solids/year. Until 1961, raw primary sludge was conditioned with ferric chloride and either kiln-dried pebble lime or spent carbide (calcium hydroxide formed as the result of chemical action in making acetylene). Bay City's conditioned sludge is dewatered on vacuum filters having an effective area of 150 sq ft. The vacuum filter cake is incinerated. In 1961, polyelectrolytes were tried as sludge conditioners in an attempt to improve filter yield. Results of filter operation for 1959-1961 using ferric chloride, lime and polyelectrolytes are presented in Table 12-4. These results clearly indicate that use of polyelectrolytes increased the filter yield and significantly reduced vacuum filtration operation time. The cost of chemicals for sludge conditioning at Bay City was $10.10/ton of dry solids when using ferric chloride and kiln-dried lime, $7.15/ton of dry solids when using ferric chloride and carbide lime and $7.00/ton of dry solids when using polyelectrolytes. The following advantages were realized when polyelectrolytes were used for sludge conditioning:

1. Savings in equipment and floor space.
2. Improved housekeeping.
3. Improved safety.
4. Reduced quantities of ash, with a large reduction in ash handling and storage.
5. Reduction in operating time, with resulting savings in operating and maintenance costs.

12.3.5 Process Design and Cost Estimates

The following example will serve to illustrate design and cost considerations when upgrading vacuum filter installations by converting from inorganic chemical to polyelectrolyte sludge conditioning.

TABLE 12-4

A COMPARISON BETWEEN LIME/FERRIC CHLORIDE AND POLYELECTROLYTES FOR CONDITIONING RAW PRIMARY SLUDGE (13)

Year	Dry Solids	Feed Solids	Filter Yield	Chemical Added, lb						Cake Solids	Solids Recovery	Operation Time
				Lime	$FeCl_3$	C-31	C-32	C-39	6-21			
	tons	percent	psf/hr							percent	percent	hr/yr
1959	[illegible]	10.2	3.0	[illegible]	[illegible]	–	–	–	–	40.1	98.2	2,125
1960	[illegible]	10.2	3.0	[illegible]	[illegible]	–	–	–	–	39.6	92.3	[illegible]
1961-62	[illegible]	10.9	5.3	–	–	–	[illegible]	–	–	35.9	72.6	[illegible]
1962-63	[illegible]	10.7	5.5	–	–	[illegible]	–	–	–	36.5	72.7	[illegible]
							Total of all three polyelectrolytes					
1963-64	[illegible]	10.2	6.3	–	–	–	←――― [illegible] ―――→			34.6	72.9	[illegible]

An existing vacuum filter installation was annually conditioning and filtering 800 tons (dry basis) of mixed digested primary and secondary sludge with a filter yield of 3 psf/hr. Sludge conditioning had been accomplished using 65 lb of ferric chloride and 300 lb of lime/ton of dry solids. To reduce the costs involved in bulk chemical handling and to increase filter yield, the applicability of a polyelectrolyte system for sludge conditioning was investigated.

The optimum polyelectrolyte dosage was found to be 20 lb/ton of dry solids, added to the digested sludge in a 1 percent solution. The polyelectrolyte addition resulted in a subsequent filter yield of 4 psf/hr. This upgrading procedure reduced the operating time of the vacuum filter by 25 percent, thereby decreasing operational and maintenance costs.

The capital cost for the polyelectrolyte application system was estimated at $7,000. This cost included all required tankage, pumps, and mixers.

12.3 Drying Beds

The dewatering of digested sludge on drying beds has long been practiced in the United States. Historically, sludge drying beds have mainly been used for smaller communities (15).

The widespread use of drying beds for smaller plants is due to their simplicity and low maintenance costs. The chemical cost and operating complexity of mechanical dewatering equipment are additional factors which favor drying beds. Disadvantages include their large land requirement, inability to dewater effectively during inclement weather, difficulty of obtaining labor for cleaning, and potential odor problems.

12.3.1 Process Considerations

One of the difficulties in developing a rational design for sludge beds is the multitude of variables which affect the sludge drying rate. In practice, it is difficult to isolate these variables and evaluate them quantitatively. Some of the more important variables are (16):

1. Climatic and atmospheric conditions
   - a. Temperature
   - b. Humidity
   - c. Rainfall
   - d. Wind velocity
   - e. Barometric pressure
   - f. Solar radiation

2. Sludge characteristics

   a. Type of sludge
   b. Method of stabilization
   c. Moisture content
   d. Grease content

3. Operational factors

   a. Sludge age
   b. Presence or absence of coagulants
   c. Depth and frequency of application

4. Bed characteristics

   a. Condition and gradation of sand
   b. Condition of drainage system
   c. Travel distance to bed extremities

Notwithstanding the magnitude of the problems involved, some generalizations concerning the applicability of these factors can be made. When possible, decisions regarding the specific effect of any or all of these factors should be based on bench scale testing.

Quon and Johnson (17) have indicated that well digested sludge should be applied to drying beds in depths of 6 to 9 inches, with 8 inches appearing to give optimum drying rates.

Sludge should be properly digested before being applied to the drying beds. Raw or poorly digested sludge dewaters slowly and produces strong odors. Sludge that has been overly digested will contain a high percentage of fine solids which will impair drainage. Aerobically digested sludge usually drains well on sand drying beds (1) (18).

Drainage and evaporation are the mechanisms which affect the dewatering rates of well-digested sludge on drying beds. It has been widely accepted that under normal conditions, practically all of the drainage of sludge occurs during the first three days following the filling of the drying bed (6). After this initial period, it was felt that evaporation was largely responsible for additional dewatering of the sludge. Recent studies indicate this is not the case (16) (17) (19). In an extended study, it was found that the initial rate of drainage was small, but that it increased with time (17). After approximately three days, the drainage rate increased and the sludge surface dropped substantially. This phenomenon is explained by considering that air trapped in the voids of the sand bed is not free to move and thus impedes the initial flow of water through the filter. Eventually, this air is liberated, allowing a greater flow to pass through the sand bed. After a period of maximum drainage, the drainage rate gradually decreases due to the buildup of solids on the sand surface, which offers resistance to further filtration. Once this point is reached, evaporation from the free water surface accounts for further dewatering.

Drying beds are usually constructed with an underdrain piping system to increase the drainage rate. Although some installations have been designed with asphalt or concrete bottoms to facilitate sludge removal, experiments have shown that drained sludge beds dry 25 percent faster than beds with an impervious bottom (17). Underdrainage from drying beds should be returned to the plant for treatment (20).

In areas with adverse climatic conditions, the use of glass covered beds, while expensive, has been found to increase the total output of dewatered sludge by 160 percent (21). Recent work in northern Texas, however, indicated that during the dry season, covers retarded the drying rate rather than accelerating it (16). Thus, where covered beds are constructed, adequate ventilation must be provided so that maximum evaporation rates may be maintained.

12.4.2 Design Basis

Present day design practices are based largely on comparisons with existing plants in the area, or upon empirical recommendations. The sludge drying bed area requirements shown in Table 12-5 are recommended for domestic wastewater treatment plants located in the northern United States (6).

TABLE 12-5
SLUDGE DRYING BED AREA REQUIREMENTS

Type of Sludge	Area of Drying Beds sq ft/capita	
	Open Beds	Covered Beds
Primary Digested	1.0 to 1.5	0.75 to 1.0
Primary and Humus Digested	1.25 to 1.75	1.0 to 1.25
Primary and Activated Digested	1.75 to 2.5	1.25 to 1.5
Primary and Chemically Precipitated Digested	2.0 to 2.5	1.25 to 1.5

In the southern United States, reduced areas are often practical because of more favorable climatic conditions.

Recently, rational design procedures have been developed for sludge drying beds (6). Equations which describe a sludge's drainage and drying properties have been derived (22). These equations lend themselves to incorporation in computer models for sludge properties, weather variability and economics. These methods may eventually replace the present empirical practice for designing drying beds.

12.3.3 Upgrading Existing Facilities

It is possible to upgrade an overloaded sludge drying bed by the following methods:

1. Improving the performance of upstream facilities, e.g., thickeners, digesters
2. Adding chemicals to improve sludge dewatering characteristics
3. Covering open beds where climatic conditions adversely affect performance.

Chemicals such as alum, ferric chloride and polyelectrolytes have been used as flocculants to improve the dewatering capacity of sludge drying beds. The use of these chemicals increases the permissible sludge loadings to the drying beds by increasing the number of bed applications per year.

In general, the chemical flocculants allow greater amounts of water to drain from the sludge, thereby decreasing the amount of water to be removed through the slower evaporation process. Bed loadings for chemically treated and untreated sludge should be evaluated by laboratory and field testing to determine the effectiveness of chemical addition on sludge drying. Buchner Funnel Tests can be used to predict dewatering rates on drying beds in the same manner as for vacuum filters (23). Care must be taken to avoid adding excess amounts of chemicals, which might blind sand particles and lower drainage rates. Alum has been used successfully at a dosage of 1 lb/100 gal of digested sludge (1). On the other hand, polyelectrolyte has been used at dosages as low as 0.025 lb/100 gal of digested sludge (16).

12.3.4 Process Designs and Cost Estimates

An open drying bed at an existing activated sludge plant was originally designed based on a population equivalent of 20,000 and a bed area requirement of 2 sq ft/capita and was loaded at a rate of 10 lb of dry solids/sq ft/yr. As a result of upgrading secondary treatment units, it was necessary to increase the loading to 15 lb/sq ft/yr to accommodate the increased sludge quantities. Two alternatives were available for upgrading the existing drying beds. It was possible either to cover the beds or to add 1 lb of alum/100 gal of digested sludge to decrease the drying time by approximately 30 percent.

Covering the drying beds was estimated to cost $240,000. The estimated cost of the alum slurry feed system and flocculation tank was $88,000.

Based on comparison of these capital costs, it appears that chemical addition may be the most economic alternative. However, to draw a definite conclusion, it would be necessary to compare total annual costs of both alternatives over the expected life of the facility.

12.4 Centrifugation

The use of centrifuges for separating materials of different densities is well established as discussed in Section 10.5. However, their use in the United States for sludge dewatering is not as widespread as the use of vacuum filters. Recent improvements in centrifuge design and the availability of full-scale performance data have encouraged their use for sludge dewatering.

12.4.1 Types of Centrifuges

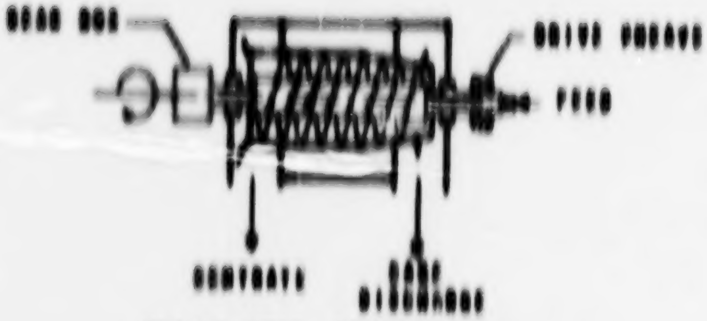
There are three general classifications of centrifuges that can be applied to sludge dewatering: solid-bowl, disc, and basket. These are illustrated on Figure 12.3. Typical centrifuge characteristics are summarized in Table 12.6.

TABLE 12.6
SUMMARY OF CENTRIFUGE CHARACTERISTICS (25)

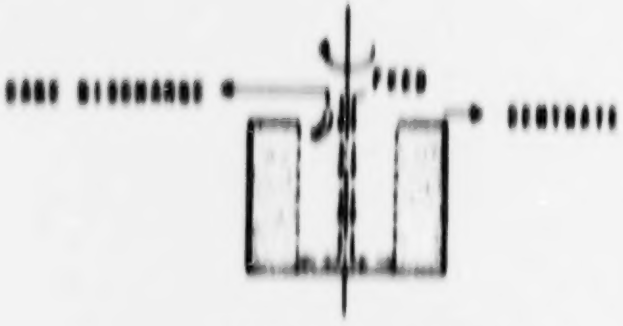
Description	Centrifuge Description		
	Solid Bowl	Basket	Disc
Method of Operation	Continuous	Batch	Continuous
Bowl Diameter, in.	6 to 60	12 to 60	8 to 30
Flow Rate, gpm	1 to 200	100 max	10 to 300
Solids in Feed, percent	*	0.1 to 30	0.1 to 10
Solids Discharged	1 to 15 tons/hr	1,000 lb max	10 to 3,000 gal/hr
Speed, rpm	1,000 to 6,000	2,500 max	4,500 to 10,000
Centrifugal Force, G	3,200 max	2,000 max	12,000 max
Motor Horsepower	5 to 150	100 max	10 to 125

*Any liquid or slurry which can be pumped.

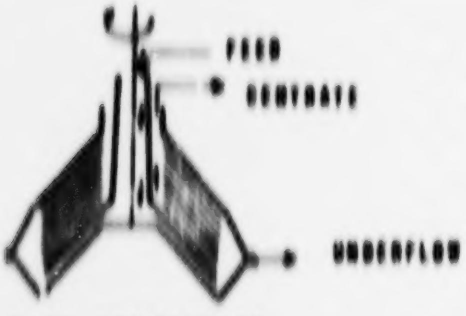
FIGURE 12-2
TYPES OF CENTRIFUGES (24)


GEAR BOX
DRIVE SHEAVE
FEED
CENTRATE
CAKE DISCHARGE

SOLID-BOWL CENTRIFUGE


CAKE DISCHARGE
FEED
CENTRATE

BASKET CENTRIFUGE


FEED
CENTRATE
UNDERFLOW

DISC CENTRIFUGE

Solid-bowl centrifuges are widely used because of their dependable performance and relatively low maintenance requirements. Solid-bowl centrifuges are suited to dewatering primary sludge and mixtures of primary and waste biological sludge. For most sludges, to achieve solids recovery in the range of 80 to 95 percent with a solid-bowl centrifuge requires the addition of polyelectrolytes.

Unlike solid-bowl centrifuges, basket centrifuges operate on a batch basis. Because of the large bowl diameter, basket centrifuges operate at slower speeds. The centrifuge can be operated on an automatic cycle for programmed filling and skimming. Basket centrifuges are generally applicable only for sludge dewatering at smaller plants.

Disc centrifuges are used primarily for thickening of waste activated sludge as discussed in Chapter 10.

12.4.2 Process Considerations

Major process variables for centrifugation are feed rate, sludge solids characteristics, feed consistency, temperature and chemical additives. Solid-bowl machine variables are bowl design, bowl speed, pool volume and conveyor speed (1). Major performance factors are cake dryness and solids recovery. To increase cake dryness, the following actions should be considered (1):

1. Increase feed rate.
2. Decrease feed solids concentration.
3. Increase temperature.
4. Do not use flocculants.
5. Increase bowl speed.
6. Decrease pool volume.
7. Decrease differential conveyor speed.

Actions which should be considered for increasing solids recovery are as follows (1):

1. Decrease feed rate.
2. Increase feed solids concentration.
3. Increase temperature.
4. Use flocculants.
5. Increase bowl speed.
6. Increase pool volume.
7. Decrease differential conveyor speed.

Most of the variables which improve cake dryness tend to decrease the solids recovery. This is an important feature of centrifuge operation. Therefore, operation of a centrifuge should be optimized to obtain the desired balance between cake dryness and solids recovery.

The following advantages are associated with solid bowl centrifugal dewatering:

1. Capital cost is less than for vacuum or pressure filters.

2. The unit is totally enclosed so that odors are more readily controlled.

3. The unit will fit in a small space and requires a minimum of auxiliary equipment.

4. A wide variety of solids can be handled.

5. Minimum operator attention is required.

Disadvantages of solid bowl centrifugal dewatering are:

1. Solids capture is often poor without the use of chemicals.

2. Chemical costs can be substantial.

3. Cake moisture is often higher and centrate quality lower than with vacuum filtration.

4. Maintenance costs are high, especially if the sludge contains substantial grit quantities.

5. Fine solids which escape in the centrate may resist settling when recycled to the head of the treatment plant, gradually build up in concentration and eventually raise effluent solids level. However, this should not be a problem if capture efficiencies can be maintained above 90 percent.

12.4.3 Design Considerations

Centrifuges should be selected on the basis of pilot tests with smaller geometrically similar machines. Proper scale-up for predictable results must consider and provide for the following variables (16):

1. Physical nature of solids being handled.

2. Stability of centrifuge feed.

3. Solids dewatering time

4. Chemical flocculant dosages

5. Percent solids recovery

6. Resistance to abrasion

7. Wet cake discharge rate.

The use of a small continuous pilot centrifuge for testing purposes may not be possible if sludge supply is limited. Several manufacturers have correlated full-scale unit performance with "spin tests" conducted on laboratory centrifuges. This technique has limited value in estimating design criteria and generally is of use only to the manufacturer.

12.4.4 Centrifuge Performance in Sludge Dewatering

Table 12-7 contains operating data for solid-bowl centrifuges for various combinations of municipal wastewater sludges.

Raw primary and digested primary sludges dewater easily; with polyelectrolyte addition, a centrifuge can produce 35 to 40 percent cake solids with better than 90 percent recovery. Waste activated sludge alone is very difficult to dewater. When waste activated or trickling filter sludge is added to raw or digested primary sludges, the cake solids content is reduced to about 20 to 25 percent, and the polyelectrolyte dosage to obtain 90 percent recovery increases. Factors responsible for this loss in efficiency include lower feed solids concentration and the poorer dewatering characteristics of biological sludge.

The use of centrifuges for sludge dewatering has been considered recently by several municipalities. At one large southeastern city, a centrifuge test program was conducted to determine the applicability of centrifugal dewatering of raw primary, digested primary, co-settled raw primary and waste activated, mixed digested and primary digested plus thickened waste activated sludges (27). With a solid-bowl centrifuge, 55 to 85 percent recovery was obtained without the use of polyelectrolytes, depending upon the feed rate. Recovery levels of 85 percent or better were achieved with 0.5 to 5.0 lb of strong cationic polyelectrolyte/ton of dry solids for combined raw primary and waste activated sludge, primary digested plus waste activated sludge and mixed digested sludge. Problems were encountered in trying to use a solid-bowl centrifuge on the combination of thickened waste activated and primary digested sludge. Higher levels of polyelectrolyte were required as the proportion of waste activated sludge to primary sludge increased. When operating with a mixture of two-thirds primary digested sludge and one-third thickened waste activated sludge on a dry solids basis, 50 to 60 percent recovery was achieved without polyelectrolytes. To increase recovery to 80 to 90 percent, polyelectrolyte dosages of 10 to 30 lb/ton of dry solids were required.

TABLE 12.7
SOLID BOWL CENTRIFUGE PERFORMANCE DATA

[illegible]

At a southwestern community, field tests were conducted to determine the applicability of centrifugation for dewatering combined primary and secondary digested sludge (37). The results indicated that a solid bowl centrifuge could be used to replace concrete drying beds. With the use of a strong cationic polyelectrolyte at a concentration of 3 to 4 lb/ton of dry solids, the sludge could be dewatered to 17 to 18 percent solids with a solids recovery of 85 percent. Solids recovery was increased to 98 to 99 percent when the polyelectrolyte dosage was raised to 5 to 6 lb/ton of dry solids.

In El Paso, Texas, a land problem necessitated replacement of sludge drying beds (12). Centrifuges were used to dewater the digested sludge at a 6 to 7 percent feed solids concentration. The centrifuges produced 30 to 32 percent cake solids with 85 to 90 percent recovery, at a polyelectrolyte dosage of 2 to 3 lb/ton of dry solids.

The use of a cationic polyelectrolyte was evaluated for a combination of 80 percent raw Imhoff and 20 percent digested sludge (37). The cationic polyelectrolyte was effective in improving sludge dewatering at economical dosage levels (2 to 3 lb/ton of dry solids). With an average feed solids concentration of 8 percent, a cake solids of 35 percent at 95 percent solids recovery was obtained.

12.4.5 Use of Centrifugal Dewatering for Upgrading Sludge Handling Facilities

The performance of centrifuges in various applications clearly indicates that centrifugation should be considered when the upgrading of sludge dewatering facilities is required. Centrifuges can be used to replace or supplement vacuum filters for sludge dewatering, especially where space is limited. Where centrifuges are considered for an upgrading situation, the efficiency of existing grit and screenings removal facilities must be carefully studied. Solid-bowl centrifuge manufacturers normally recommend that cyclonic grit removal be provided ahead of the centrifuge to reduce abrasion. Inadequate removal or comminution of screenings can result in plugging of the centrifuge inlet. As centrifuges must be periodically dismantled for maintenance, provision for an overhead hoist is desirable.

12.4.6 Process Design and Cost Estimates

The following example is presented to illustrate the upgrading of sludge handling facilities through the use of centrifuges.

Vacuum filter facilities at an existing activated sludge plant were overloaded due to an increase in plant flow from 10 to 30 mgd. Space limitations at the plant prohibited the installation of additional vacuum filters.

It was decided to add centrifugal dewatering to handle the additional 12,000 lb/day of dry solids (30 gpm by volume). Facilities for sludge dewatering included two solid-bowl centrifuges, each capable of handling 15 to 30 gpm, sludge feed pumps, polyelectrolyte addition facilities and other necessary appurtenances. The solid-bowl centrifuges were designed to produce 20 to 25 percent cake solids at 85 percent recovery, with 3 percent feed solids and a polyelectrolyte dosage of 3 to 6 lb/ton of dry solids. The capital cost for this upgrading procedure was estimated at $296,400.

12.8 Filter Presses

A variety of filter presses have been used to dewater sludge. The most commonly used type consists of a series of rectangular plates, recessed on both sides, and supported face to face in a vertical position on a frame with a fixed and movable head. A filter cloth is hung or fitted over each plate. The plates are held together with sufficient force to withstand the pressure applied during the filtration process. Hydraulic rams or powered screws are used to hold the plates together.

In operation, chemically conditioned sludge is pumped into the space between the plates at a pressure of 60 to 180 psig. Under this pressure, the liquid is forced through the filter cloth and plate outlet ports. The plates are then separated and the sludge drops to a hopper below. Filtrate normally is returned to the headworks of the treatment plant. The sludge cake thickness varies from 1 to 1 1/2 inches and the moisture content from 55 to 70 percent. A complete filtration cycle includes time required for pressing, cake removal, media washing and press closing. Total filtration time varies from three to six hours with pressing time ranging from one to three hours.

The major operating costs associated with this method of dewatering are those for chemical conditioning and maintenance and replacement of filter cloths. Although not commonly used in the United States, filter presses have been used extensively in Europe.

The advantages of filter presses are as follows:

1. Low cake moisture permits incineration of primary/secondary sludge combinations without auxiliary fuel.

2. Continuous operating attendance is less than that required for vacuum filters.

3. A large filtration area can be installed in a minimum of floor area.

4. Hard to dewater sludges can be handled more readily.

Major disadvantages are as follows:

1. Capital costs are considerably higher than for centrifuges or vacuum filters.

2. Batch discharge requires equalization of pressed cake production prior to incineration.

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CHAPTER 13

CASE HISTORIES OF TREATMENT PLANT UPGRADING

13.1 Case History No. 1 - Use of Roughing Filter to Upgrade an Existing Low Rate Trickling Filter Plant (1) (2) (3)

Case History No. 1 involves the upgrading of an existing low rate trickling filter plant in Huber Heights, Ohio. The original plant was designed in August, 1956, for a flow of 0.7 mgd, with 85 percent BOD and SS removals. The community developed so rapidly that by 1970, the average flow had increased to 2.3 mgd. The flow diagrams for the original and upgraded plants are shown on Figure 13-1.

Operating and performance data for the overloaded plant for 1962, when the plant was receiving 1.15 mgd, are compared with corresponding data after the plant was upgraded to 2.3 mgd in Table 13-1.

The comminutor and primary clarifiers were replaced with three Hydrasieve units of 1 mgd capacity each. These units are stationary screens capable of removing 20 to 35 percent of the BOD and SS in raw wastewater. A Hydrasieve unit is illustrated on Figure 13-2, along with a flow diagram through the unit. These screens generally require no power and little maintenance; however, a head loss does occur across the units.

The plastic media roughing filter used in the upgrading operates at an application rate of approximately 2.5 gpm/sq ft. Present BOD removal is about 25 to 35 percent across the roughing unit. Because of the increased hydraulic loading, it was necessary to expand the secondary clarification and chlorine contact tank capacities. The abandoned primary clarifiers were converted to sludge thickeners. This step, in addition to conversion of the low rate anaerobic digester to a high rate unit using gas recirculation for mixing, enabled the sludge handling system to process the increased quantity of sludge produced.

This case history emphasizes that an existing plant may be gradually upgraded to handle a three-fold increase in flow with the use of innovative techniques and newly applied process equipment. The capital costs for upgrading the capacity of the plant were estimated at approximately $152,000.

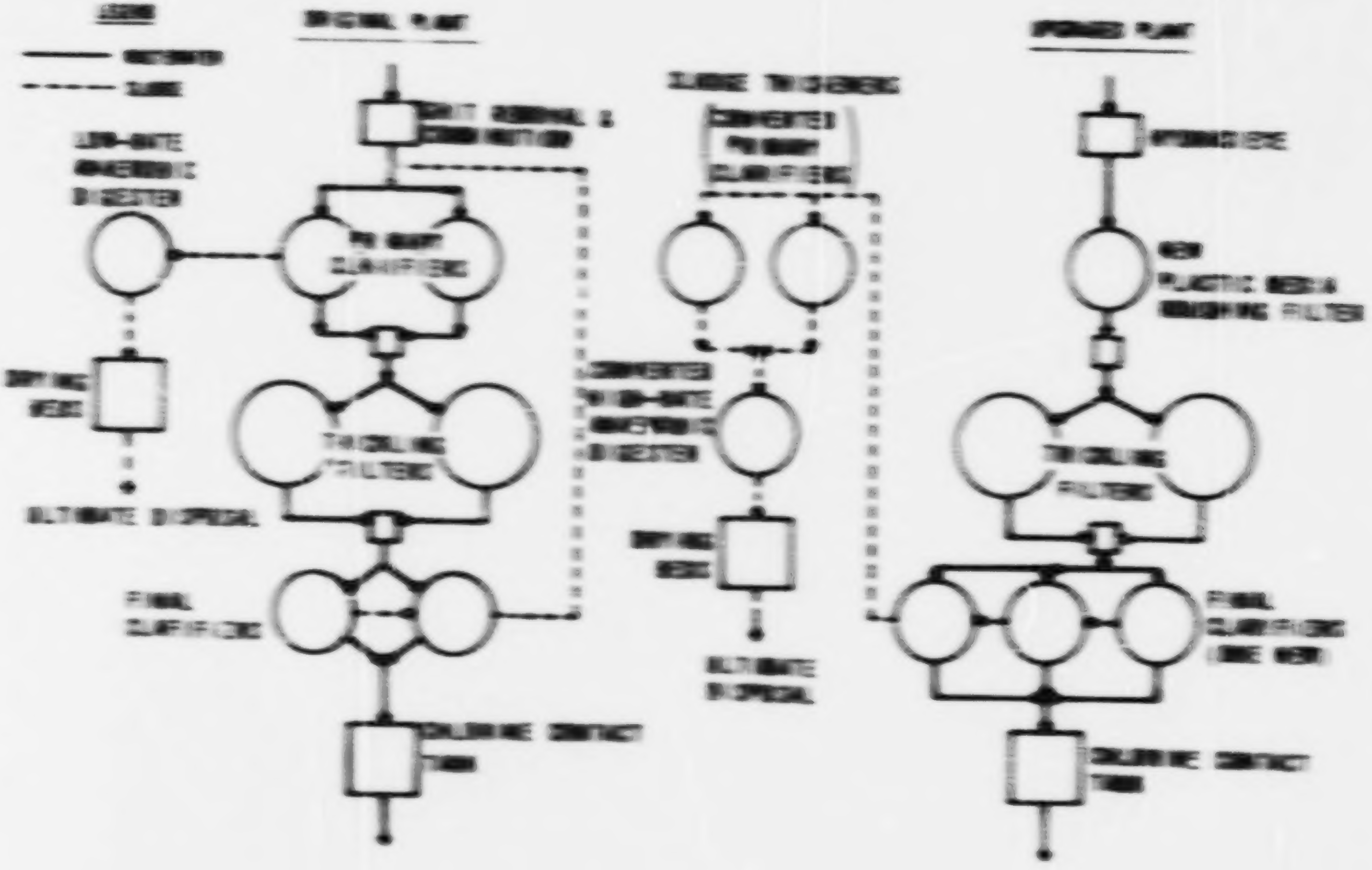
13.2 Case History No. 2 - Upgrading an Existing High Rate Trickling Filter by Conversion to a Super Rate Filter System (4)

The North Treatment Plant at Sedalia, Missouri, a high rate trickling filter plant designed for 1.25 mgd, was achieving 85 percent BOD removal in 1963. However, the State of Missouri

FIGURE 13-2

CASE HISTORY NO. 2

COMPARISON OF ORIGINAL AND UPGRADED FLOW DIAGRAMS



LEGEND
WASTEWATER
SLUDGE
ORIGINAL PLANT
GRIT REMOVAL & COMMINUTION
LOW-RATE ANAEROBIC DIGESTER
PRIMARY CLARIFIERS
DRYING BEDS
TRICKLING FILTERS
ULTIMATE DISPOSAL
FINAL CLARIFIERS
CHLORINE CONTACT TANK
SLUDGE THICKENERS
(CONVERTED PRIMARY CLARIFIERS)
CONVERTED HIGH-RATE ANAEROBIC DIGESTER
DRYING BEDS
ULTIMATE DISPOSAL
UPGRADED PLANT
HYDROSIEVE
NEW PLASTIC MEDIA ROUGHING FILTER
TRICKLING FILTERS
FINAL CLARIFIERS (ONE NEW)
CHLORINE CONTACT TANK

TABLE 11-1
CASE HISTORY NO. 1 - PLANT OPERATING AND PERFORMANCE DATA

Description	Before Upgrading (1962)	After Upgrading
Average Flow Rate, mgd	1.15	2.3
Primary Clarifier		[2]
Overflow Rate[1], gpd/sq ft	1,170	
Hydrasieve Slot Size, in		0.06
BOD Removal, percent	35[3]	25[4]
SS Removal, percent	[5]	33[4]
Plastic Media Roughing Filter		
Hydraulic Loading[6], gpm/sq ft		2.5
Organic Loading[6], lb BOD/day/1,000 cu ft		530
Recirculation Ratio		≈ 2.0
BOD Removal, percent		40
Trickling Filter (Stone Media)		
Hydraulic Loading, mgd/acre	6.0	12.0
Organic Loading, lb BOD/day/1,000 cu ft	56.2	87.0
Final Clarifier		
Overflow Rate, gpd/sq ft	1,170	750
Overall Plant Performance		
BOD Removal, percent	83	85
SS Removal, percent	[5]	84
Effluent BOD, mg/l	33	[illegible]
Effluent SS, mg/l	[5]	40

[1] Based on average flow rate

[2] Primary clarifier converted to gravity thickener

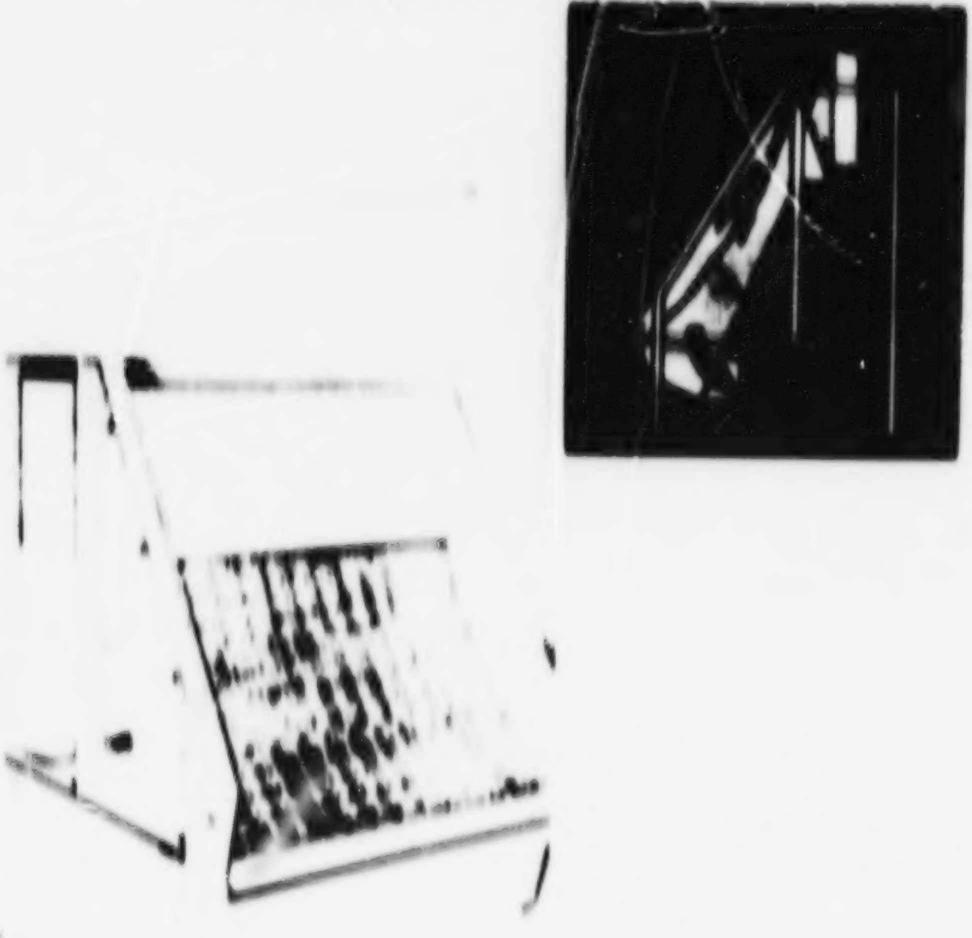
[3] Based on primary clarifier performance

[4] Based on hydrasieve performance only

[5] Performance data not available

[6] Including recirculation

FIGURE 13-2
HYDRASIEVE SCREENING UNIT *



*COURTESY OF THE BAUER BROS. CO. - SPRINGFIELD, OHIO

Water Pollution Board set a new final effluent BOD requirement of 20 mg/l, which Sedalia could not meet with the existing facilities. The 1963 plant flow diagram is illustrated on Figure 13.3.

The plant was upgraded to treat an average design flow of 2.5 mgd. The existing stone media filter was renovated to operate with plastic media. In addition, a second plastic media filter was constructed. The two plastic-media filters are operated in parallel with a total recirculation ratio of 1.55. One additional primary clarifier and one additional secondary clarifier were installed.

To remove additional BOD and SS, a shallow aerobic polishing lagoon was constructed after the secondary clarifiers, and a vacuum filter was added to dewater the additional volume of digested sludge. A flow diagram of the upgraded plant is also shown on Figure 13.3. Table 13.2 contains a summary of operating and performance data for the 1963 overloaded period, along with the upgraded design criteria and actual operating and performance data for the post-upgrading period. It should be noted that the effluent BOD was improved from 113 mg/l to 11 mg/l after upgrading, which is considerably below the 20 mg/l requirement.

The capital costs of upgrading the plant were estimated at $3,055,000.

13.4 Case History No. 3 - Upgrading Using Polyelectrolyte Addition Before the Primary Clarifier (5)

The Easterly Wastewater Treatment Works in Cleveland, Ohio, is a conventional activated sludge plant whose dry weather design flow is 123 mgd. In 1968, the plant was hydraulically overloaded. Operating data from the 1968 control period are presented in Table 13.3. It is interesting to note that the overflow rate of the primary clarifiers was 2,050 gpd/sq ft, which is substantially above normally accepted values. As a result of the hydraulic overloading, the overall BOD and SS removals were only 79 and 85 percent, respectively.

To improve the overall plant performance, a polyelectrolyte addition program was initiated. An anionic polyelectrolyte, Purifloc A-23, was added at an average dosage of 0.21 mg/l. Since proper polyelectrolyte dispersal and uniform mixing into the entire waste flow constitute an extremely important aspect of the flocculation process, it was decided to add the polyelectrolyte at the plant's two Venturi meters. These meters are located immediately after the grit chamber and in front of the preaeration basin. Dye studies indicated that there was a 7.5-minute travel time between the Venturi meters and the primary clarifiers. Six of the 7.5 minutes were spent in the preaeration basin. The gentle agitation in the preaeration basin provided adequate flocculation of wastewater solids.

FIGURE 13-3

CASE HISTORY NO. 2

COMPARISON OF ORIGINAL AND UPGRADED FLOW DIAGRAMS

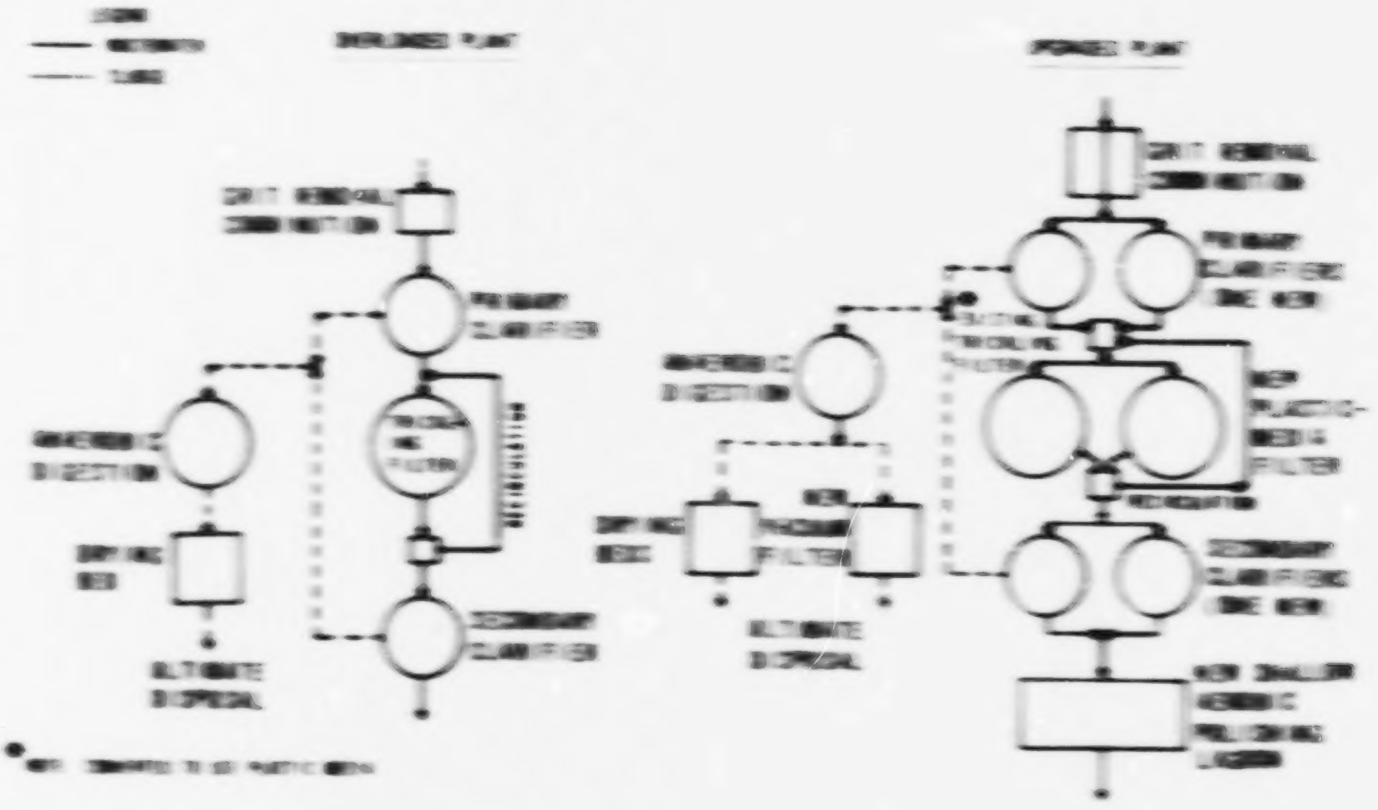

LEGEND
WASTEWATER
SLUDGE
OVERLOADED PLANT
UPGRADED PLANT
GRIT REMOVAL COMMINUTION
PRIMARY CLARIFIER
TRICKLING FILTER
RECIRCULATION
SECONDARY CLARIFIER
ANAEROBIC DIGESTION
DRYING BED
ULTIMATE DISPOSAL
GRIT REMOVAL COMMINUTION
PRIMARY CLARIFIERS (ONE NEW)
EXISTING TRICKLING FILTER
NEW PLASTIC-MEDIA FILTER
RECIRCULATION
SECONDARY CLARIFIERS (ONE NEW)
NEW SHALLOW AEROBIC POLISHING LAGOON
ANAEROBIC DIGESTION
DRYING BEDS
NEW VACUUM FILTER
ULTIMATE DISPOSAL
NOT CONVERTED TO 1ST PLASTIC MEDIA

TABLE 11-2
CASE HISTORY NO. 2 - PLANT OPERATING, PERFORMANCE AND DESIGN DATA

Description	1963 Operating and Performance Data	Upgraded Design	1969 Operating and Performance Data
Average Daily Flow, mgd	1.25	2.5	1.800
Raw Wastewater BOD, mg/l	760	570	480
Primary Clarifier			
Overflow Rate¹, gpd/sq ft	1,000	1,000	720
BOD Removal, percent	40	40	60
Trickling Filter			
Hydraulic Loading², mgd/acre	25	12	23
Organic Loading², lb BOD/day/1,000 cu ft	226	76	40
Recirculation Ratio	1.0	1.55	1.55
Final Clarifier			
Overflow Rate¹, gpd/sq ft	755	755	545
Secondary BOD Removal, percent	75	93.2	86.8
Polishing Lagoon (Shallow Aerated)			
Maximum BOD Loading, lb BOD/acre/day		60	40
BOD Removal, percent		12	54
Vacuum Filtration Rate, lb/sq ft/hr		5.0	³
Overall Plant Performance			
BOD Removal, percent	85	96.5	97.7
Effluent BOD, mg/l	115	20	11

[1] Based on average daily flow

[2] Including recirculation

[3] Lack of sufficient operating data

TABLE 13-3
CASE HISTORY NO. 3 PLANT OPERATING AND PERFORMANCE DATA

Description	1968 Control Period	Polyelectrolyte Demonstration Period
Influent BOD, mg/l	104	67
Influent SS, mg/l	169	157
Primary Clarifier		
Overflow Rate, gpd/sq ft	2,030	2,170
BOD Removal, percent	31	46
SS Removal, percent	31	51
Sludge Solids Concentration, percent	4.1	4.3
Sludge Volume Pumped, million gallons/month	5.0	6.8
Aeration Tank		
MLSS, mg/l	1,670	1,602
Organic Loading, lb BOD/day/lb MLSS	0.48	0.29
Dissolved Oxygen, mg/l	4.2	4.8
Waste Activated Sludge Concentration, percent	2.4	2.0
Waste Activated Sludge Pumped, million gallons/month	12.3	9.8
Overall Plant Performance		
BOD Removal, percent	79.1	83.4
SS Removal, percent	85.4	89.2
Effluent BOD, mg/l	21.8	11.1
Effluent SS, mg/l	24.6	17.0

A summary of the effectiveness of the polyelectrolyte addition on plant performance as compared to performance during the previous control period is also presented in Table 13-3. The improvement in primary SS removal increased the volume of primary sludge from 5.0 to 6.8 million gallons per month. Overall plant performance was improved, as noted by the reduced effluent BOD and SS concentrations. In addition to the increased treatment efficiency, the polyelectrolyte addition was responsible for the following benefits to the downstream activated sludge process:

1. A 20 percent volume decrease in waste activated sludge production.

2. A 22 percent reduction in air supply requirements, resulting in a power cost savings of over $3,300 per month.

3. An increase in aeration tank DO concentration from an average of 3.2 mg/l to 5.0 mg/l.

An economic comparison was made between polyelectrolyte addition and providing additional tankage to equal the performance of the flocculation system. The amortized cost for the additional tankage was about $369,000 per year, while the chemical cost was $150,000 per year, thereby indicating a considerable yearly savings in favor of the polyelectrolyte alternative.

13.4 Case History No. 4 - Upgrading a Trickling Filter Plant by Adding Activated Sludge Treatment and Pre- and Post-Chlorination

In 1967, the Livermore, California Wastewater Treatment Plant was upgraded to increase plant capacity from 2.5 to 5.0 mgd and to provide a higher degree of treatment (6). All of the existing treatment facilities were used in the upgraded treatment process. Flow diagrams that show the type, number and arrangement of process units before and after upgrading are presented on Figure 13-4.

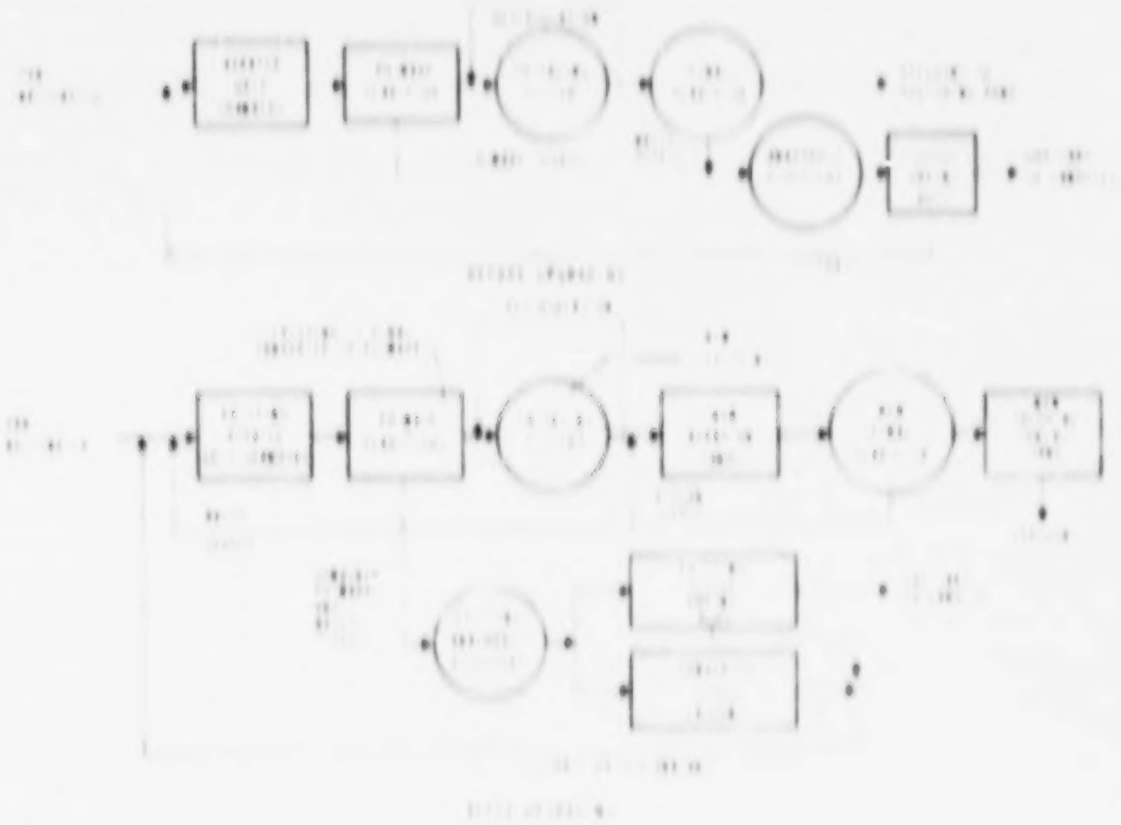
The original plant was constructed in 1958 to provide secondary treatment for a domestic wastewater flow of 2.5 mgd. Effluent standards in effect at that time required that the plant produce an effluent that contained no more than 40 mg/l of BOD and 40 mg/l of SS. To achieve this degree of treatment, preliminary and primary treatment facilities, a stone media trickling filter, a final clarifier and a 30-day polishing pond were provided. Waste sludge was stabilized by anaerobic digestion and dewatered on sludge drying beds. This treatment process produced an effluent that contained 45 to 50 mg/l of BOD and 45 to 50 mg/l of SS.

The upgraded plant was designed to produce an effluent that contained not more than 20 mg/l of BOD, 20 mg/l of SS and 1 mg/l of grease. Bacteriological requirements limited the average total coliform count to not more than 5 MPN per 100 ml over a five-day period. To meet these design criteria, the plant facilities were expanded to provide activated sludge treatment following first-stage trickling filtration along with pre- and post-chlorination. As indicated on Figure 13-4, the existing grit chambers were adequate for the increased flow. Sufficient primary clarification capacity at the higher flow was provided by converting the existing final clarifier to a primary clarifier. An additional stone media trickling filter was constructed to provide for the increased hydraulic loading. To improve the degree of treatment, two new aeration tanks, a new larger final clarifier and a new chlorination contact tank were added to the treatment sequence. The polishing pond was converted to an emergency holding basin from which any substandard effluent could be recycled through the treatment plant. The sludge handling and treatment system was designed to return excess waste activated sludge to the head end of the plant, and to anaerobically digest combined primary and waste activated sludge. The existing sludge drying beds and two new sludge lagoons were used for dewatering the digested sludge. A cost saving feature of the upgraded plant is that trickling filter effluent is fed directly to the aeration tanks without intermediate clarification.

FIGURE 13-4

CASE HISTORY NO 4

COMPARISON OF ORIGINAL AND UPGRADED FLOW DIAGRAMS



Design conditions before and after upgrading are presented in Table 13-4. A summary of treatment performance before and after upgrading is presented in Table 13-5.

As indicated in Table 13-5, the plant is now producing an effluent that well exceeds the effluent standards for which it was designed. The effluent concentrations of BOD and SS represent, respectively, 96.6 and 94.3 percent removals of the raw wastewater concentrations of these two constituents.

The plant is presently achieving 99 percent oxidation of ammonia nitrogen. This is important, not only because of the oxygen demand associated with ammonia nitrogen, but also because the presence of ammonia compounds would require excessive chlorine dosages to achieve the required high levels of disinfection.

The plan for upgrading requires that the existing digesters operate as high-rate units to accommodate the increased quantity of sludge. Under present operating conditions, each digester is loaded at an average rate of 0.22 lb of dry solids/cu ft/day. At this loading rate, approximately 54 percent reduction in volatile solids is achieved and the digested sludge contains 1.6 percent solids by weight. This low concentration of solids coupled with the lack of mechanical cleaning facilities has caused some difficulty in attaining optimum use of the sludge drying beds. Accordingly, most of the digested sludge is dewatered in the sludge lagoons. The recycled flow from the sludge lagoons has caused some difficulties in plant operation. For these reasons, various plans for upgrading the sludge treatment and disposal system are now under consideration (7).

The cost of upgrading the plant was $1,930,000 and the annual operating costs for 1970 were estimated at $227,840, exclusive of administrative costs.

13.5 Case History No. 5 - Upgrading a Primary Treatment Plant to Provide Tertiary Treatment (8)

The Central Contra Costa Sanitary District of Walnut Creek, California, undertook plant-scale studies to determine the best method of upgrading a primary plant to produce an effluent that would be acceptable for industrial cooling water use. The findings of this investigation indicated that if the existing plant facilities were modified and expanded to provide carbonaceous and nitrogen oxidation, denitrification and filtration, the effluent would be acceptable for the intended use. Flow diagrams for the existing and proposed plant are shown on Figure 13-5.

Operating conditions for the existing plant and proposed design parameters for the upgraded plant are shown in Table 13-6. Most of the proposed design parameters are based on the results obtained from the plant-scale studies.

TABLE 13-4
CASE HISTORY NO. 4 - PLANT OPERATING AND DESIGN CONDITIONS

Description	Operating Conditions Before Upgrading	Design Conditions After Upgrading
Pretreatment Facilities Capacity, mgd	3.5	5.0
Preaeration Tanks		
Number of Units	2	2
Detention Time, hr	0.6[1]	0.6[2]
Air Supplied per tank, cfm	300	150
Hydraulic Capacity, mgd	10	10
Primary Clarifiers		
Number of Units	2	2
Overflow Rate, gpd/sq ft[1]	1,050	1,050
Hydraulic Capacity	10	10
Trickling Filters		
Number of Units	2	2
Hydraulic Loading, mgd/acre	34.5	34.5
Organic Loading, lb BOD/day/acre ft	4,600	4,400
Recirculation Ratio	1.5 to 3.0	1.5 to 3.0
Aeration Tanks		
Number of Units	—	2
Detention Time, hr	—	5.2
Volumetric Loading, lb BOD/day/1,000 cu ft	—	20
Sludge Recycle, percent[1]	—	10 to 100
Final Clarifier		
Number of Units	1	1
Detention Time, hr	3.0	2.75
Overflow Rate, gpd/sq ft[1]	822	787
Polishing Pond		
Number of Units	1	—
Detention Time, days	40	—
Organic Loading, lb BOD/day/acre ft	30	—

TABLE 11-4 (Continued)

Description	Operating Conditions Before Upgrading	Design Conditions After Upgrading
Anaerobic Digesters		
Number of Units	2	2
Loading, lb dry solids/cu ft/day	0.07	0.22
Chlorine Contact Tank		
Number of Units		1
Detention Time, hr		1.0
Sludge Drying Beds		
Number of Units	4	4
Total Area, sq ft	22,400	22,400
Digested Sludge Lagoons		
Number of Units		2
Volume each, 1,000 cu ft		120

[1] Only one tank in service.
[2] Both tanks in service.
[3] Average dry weather flow.

TABLE 11-5
SUMMARY OF TREATMENT PERFORMANCE FOR CASE HISTORY NO. 4

Parameter	Measured Performance Before Upgrading	Measured Performance After Upgrading[2]
Average Flow, mgd	2.5	3.47
Effluent BOD, mg/l	15-30	7.4
Effluent SS, mg/l	15-30	11
NH_3-N, mg/l	1	0.11
NO_3-N, mg/l	1	21.5
Coliforms, MPN/100 ml		2.5
Grease, mg/l		0.23

[1] Before upgrading, the plant produced no significant nitrification.
[2] Monthly average.

FIGURE 13-5

CASE HISTORY NO.5

COMPARISON OF ORIGINAL AND UPGRADED FLOW DIAGRAMS

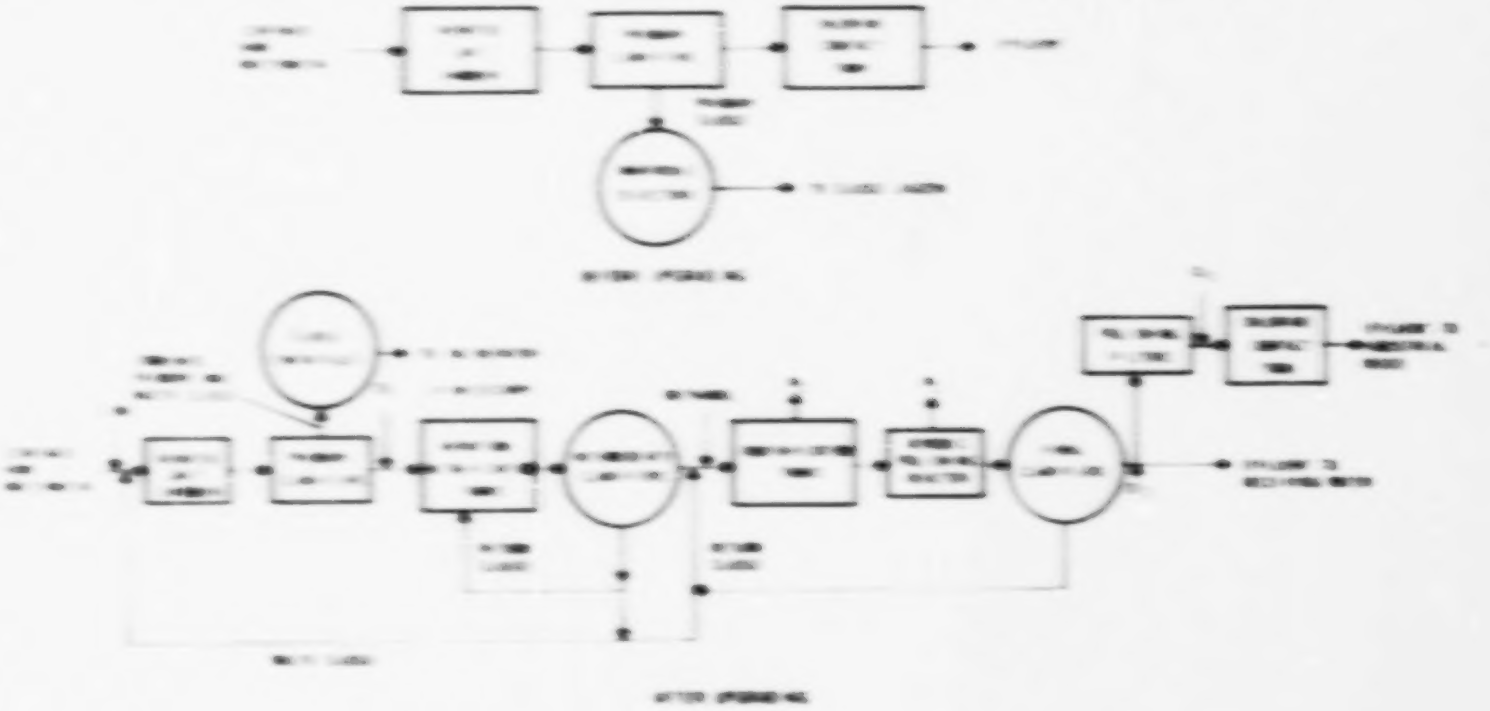


TABLE 18-6
CASE HISTORY NO. 5 - PLANT OPERATING AND DESIGN CONDITIONS

Description	Operating Conditions Before Upgrading	Design Conditions After Upgrading
Preliminary Treatment (Bar Screen, Raw Wastewater Pumps) Average Capacity, mgd	31	30
Preaeration Grit Removal Tanks		
Detention Time (ADWF[1]), hr	0.5	0.37
Design Flow, mgd	31	30
Primary Clarifiers		
Number of Units	7	4
Detention Time (ADWF), hr	1.5 - 1.6[2]	2.4
Overflow Rate (ADWF), gpd/sq ft	1,040 - 1,200	720
Aeration and Nitrification Tanks		
Number of Units		2
Detention Time, hr		6.8
Volumetric Loading, lb BOD/day/1,000 cu ft		27.3
Intermediate Clarifiers		
Number of Units		4
Detention Time (ADWF), hr		3
Overflow Rate (ADWF), gpd/sq ft		700
Overflow Rate (peak flow), gpd/sq ft		2,000
Denitrification Tanks		
Number of Units		2
Detention Time (ADWF), minutes		104
Final Clarifiers		
Number of Units		4
Detention Time (ADWF), hr		3
Overflow Rate (ADWF), gpd/sq ft		700
Overflow Rate (peak flow), gpd/sq ft		2,000
Effluent Polishing Filters		
Number of Units		4
Filtration Rate (ADWF), gpm/sq ft		4.0
Backwash Water Rate, gpm/sq ft		
Maximum Rate		25
Minimum Rate		10

TABLE 11-6 (Continued)

Description	Operating Conditions Before Upgrading	Design Conditions After Upgrading
Sludge Digesters		
Number of Units	4	
SRT, days	25	
Loading, lb dry solids/cu ft/day	0.15	
Sludge Dewatering and Incineration		
Centrifuges, Number of Units		2
Rate, gpm		100
Polymer dosage, lb/ton of dry solids		4 - 5
Incinerators, Number of Units		1
Furnace Capacity, lb/hr		31,200
Sludge Drying Beds		
Number of Units	4	
Total Area, sq ft	20,000	
Sludge Lagoons		
Number of Units	3	
Capacity, 1,000 lb dry solids	15,600	
Chlorination, lb/day	10,000	

[1] ADWF = Average dry weather flow

[2] Varies from tank to tank

The original primary treatment plant was constructed in 1948, with expansions in 1957 and 1964 to increase the plant capacity to 33 mgd. Pretreatment including pre-chlorination, screening, and grit removal, followed by primary sedimentation and chlorine disinfection produced an effluent with 144 mg/l of BOD. Sludge produced in the primary tanks was anaerobically digested and then discharged to sludge lagoons and drying beds.

The upgraded plant is designed to produce an effluent that is suitable for either industrial reuse or discharge into receiving waters. As indicated on Figure 11-5, the design calls for the addition of lime to the wastewater flow, after screening. The plant-scale field studies indicated that when lime is mixed with the wastewater in the aerated grit chambers, it combines with phosphorus and other substances to form a floc that improves the coagulation of solids. As a result, increased BOD, SS and phosphorus removals may be expected in the primary clarifiers. Effluent from the primary tanks will be pumped to the aeration-nitrification tanks, where carbonaceous organics and ammonia nitrogen will be simultaneously oxidized. The increased BOD removal accomplished in the primary clarifiers due to lime addition lowers the organic loading on the aeration tanks and is the key upgrading feature that will permit nitrification in a single-stage activated sludge system. The previously mentioned plant-scale studies also indicated that recarbonation of the lime-treated primary effluent prior to entering the aeration tanks was not necessary when the lime clarification process was operated at a pH of 11.0 or less, due to the CO_2 and nitric acid produced in the nitrification reactor.

Nitrified effluent from the intermediate clarifiers will be fed to a stirred anaerobic reactor for biological denitrification. Methanol will be added as an energy source for the denitrifying organisms. Following denitrification, the wastewater passes through a short detention aerobic polishing reactor for oxidizing residual methanol and for stripping nitrogen gas formed prior to final clarification. After final clarification, the effluent will be chlorinated and the portion required for industrial use pumped to dual-media polishing filters and final chlorine disinfection. Any remaining effluent will be discharged to the receiving water.

The sludge handling and treatment system will be upgraded and expanded to accommodate the increased quantities of sludge. Waste biological sludge will be returned to the plant influent ahead of the point of lime addition and settled in the primary tanks together with the chemical-primary sludge. The existing sludge digesters will be modified to serve as holding and mixing tanks for the combined sludge. The blended sludge will be classified and dewatered by two-stage centrifugation prior to incineration in multiple hearth furnaces.

The upgraded plant will provide for an average dry weather flow capacity of 45 mgd through the primary treatment units and 30 mgd through the remainder of the plant. The design and layout for this upgrading were planned such that subsequent plant expansions will ultimately provide the capacity to treat a future average dry weather flow of 120 mgd. All hydraulic structures for the primary and secondary treatment facilities are designed to carry the ultimate maximum wet weather flow of 300 mgd.

Design criteria for the existing and the upgraded plant are presented in Table 13-6. A summary of the treatment plant performance before upgrading and the anticipated performance after upgrading is given in Table 13-7. Costs for the upgrading are presented in Table 13-8.

TABLE 14-7
SUMMARY OF TREATMENT PERFORMANCE FOR CASE HISTORY NO. 5

Parameter	Measured Performance Before Upgrading	Anticipated Performance After Upgrading[1]
Average Design Flow, mgd	31	40
Effluent BOD, mg/l	103 – 147	3
Effluent SS, mg/l	57 – 64	1
Effluent Total Nitrogen, mg/l		2
Effluent Total Phosphorus, mg/l		0.2

[1] after effluent polishing for industrial reuse

TABLE 14-8
ESTIMATED COSTS FOR CONSTRUCTION FOR CASE HISTORY NO. 5

Description	Estimated Cost
Preliminary Treatment Works	$ 600,000
Preaeration Grit Removal Structure	540,000
Primary Clarifiers	920,000
Aeration and Nitrification Tanks	4,400,000
Aeration Blower Building and Equipment	1,620,000
Intermediate Clarifiers	1,710,000
Denitrification Tanks	860,000
Final Clarifiers	1,670,000
Primary Effluent and Final Effluent Pumping Structure	1,130,000
Incineration and Dewatering Building and Equipment	4,000,000
Chemical Storage Area	650,000
Dual Media Filters	1,900,000
Effluent Storage, Chlorination and Pumping Facilities	2,300,000
Administration Building Laboratory	1,130,000
Maintenance Building	610,000
Utility Tunnel	950,000
Outside Piping and Site Development	1,400,000
Total Construction Cost	$29,110,000

13.6 Case History No. 6 - Upgrading a Trickling Filter Plant in Stages to an Activated Sludge Plant with Roughing Filters (6)

The original South Buffalo Creek Wastewater Treatment Plant in Greensboro, North Carolina, was built in 1931 for an average flow of 4.25 mgd. This facility provided primary clarification and secondary treatment using fixed-nozzle square tank trickling filters. Sludge was digested anaerobically and dewatered on sludge drying beds.

The existing plant was first upgraded in 1957 when treatment efficiency became inadequate due to an increase in average flow to 8.0 mgd, including 1.5 mgd of industrial wastes. The upgrading included replacement of the screening and grit removal units, expansion of primary clarification, final clarification and digestion facilities, installation of new circular trickling filters and addition of chlorine disinfection facilities. Replacement of the drying beds with a vacuum filter installation was also required due to the additional sludge produced by the upgraded treatment system. The treatment provided by this upgraded system resulted in BOD and SS removals of 89 percent and 84 percent, respectively.

Increases in the industrial BOD influent loads beyond those anticipated in the 1957 design required further plant upgradings in 1964 and 1970. In each of these upgradings, one of the two original fixed-nozzle trickling filters was converted into two aeration basins for activated sludge treatment. Waste activated sludge was returned to the plant influent and settled with the primary sludge. In 1970, digestion of the combined sludge was discontinued and trucking of the dewatered raw sludge to an incinerator replaced landfilling for ultimate disposal.

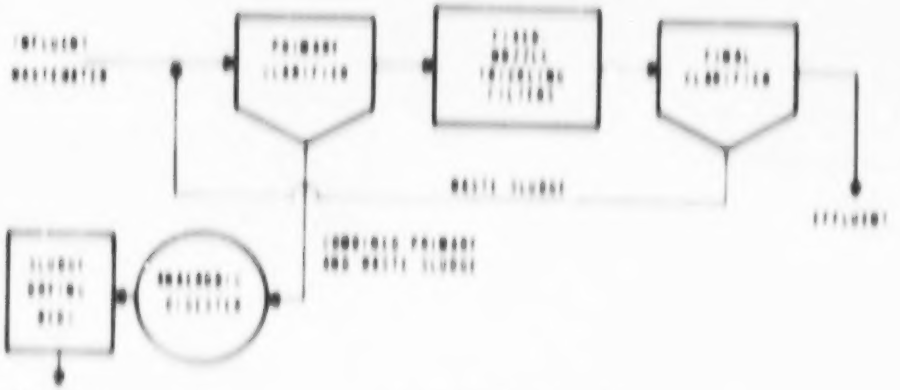
Current upgrading (presently under construction) to increase plant capacity will include chemical addition to the raw wastewater for phosphorus removal, separate flotation thickening of waste activated sludge to reduce the loading on the dewatering facilities and effluent polishing including postaeration and dual-media filtration. Under this current upgrading, the plant will provide 90 percent BOD and SS removals for an ultimate dry weather flow of 16.7 mgd.

A flow diagram and design data for the original plant and each subsequent upgrading are shown on Figure 13-6 and Table 13-9, respectively. Plant performance data before and after each upgrading are shown in Table 13-10.

Capital costs for each stage of the upgrading are presented in Table 13-11.

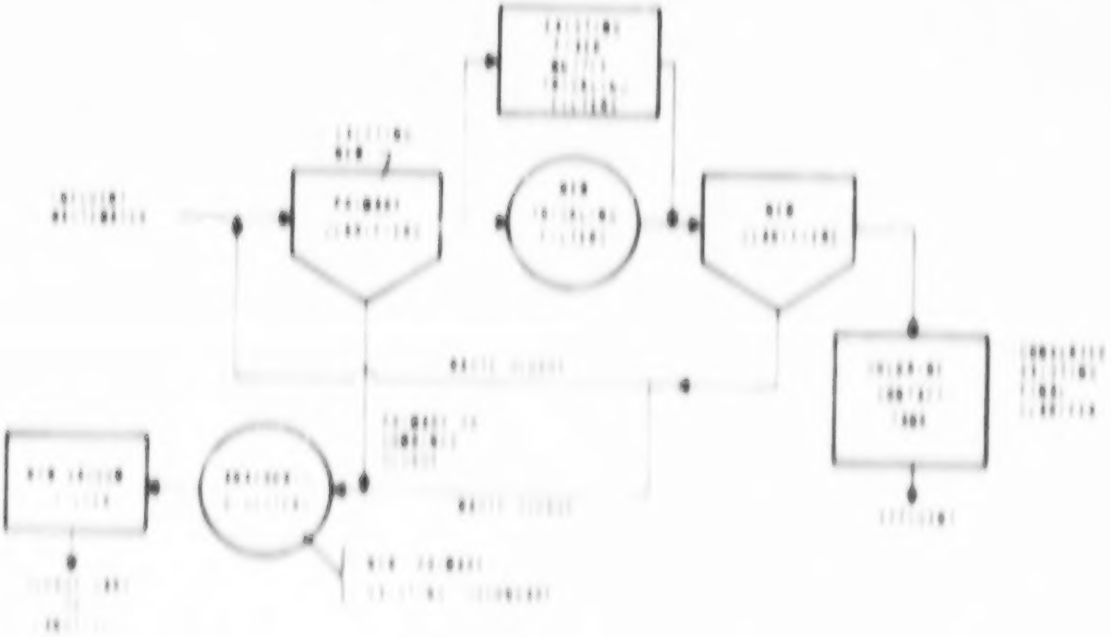
FIGURE 13.6

COMPARISON OF ORIGINAL AND UPGRADED FLOW DIAGRAMS



INFLUENT WASTEWATER
PRIMARY CLARIFIER
FIXED NOZZLE TRICKLING FILTERS
FINAL CLARIFIER
WASTE SLUDGE
EFFLUENT
COMBINED PRIMARY AND WASTE SLUDGE
SLUDGE DRYING BEDS
ANAEROBIC DIGESTER
TO LANDFILL

ORIGINAL PLANT [illegible]



INFLUENT WASTEWATER
EXISTING
NEW
PRIMARY CLARIFIERS
EXISTING FIXED NOZZLE TRICKLING FILTERS
NEW TRICKLING FILTER
NEW CLARIFIERS
WASTE SLUDGE
CHLORINE CONTACT TANK
CONVERTED EXISTING FINAL CLARIFIER
EFFLUENT
WASTE SLUDGE
NEW PRIMARY
EXISTING SECONDARY

PLANT AFTER [illegible] UPGRADING

FIGURE 13.6

(CONTINUED)

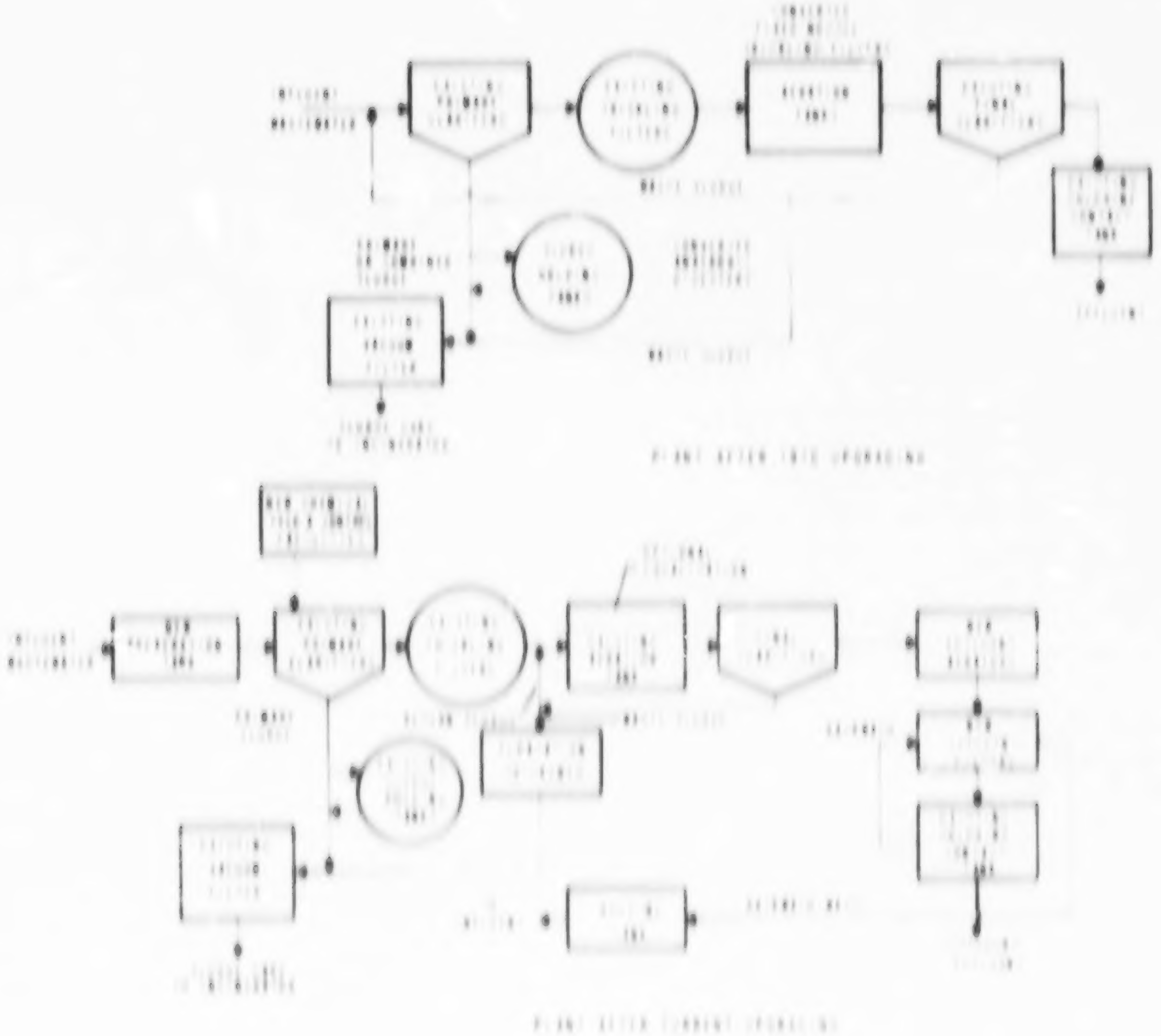


TABLE 13-9
CASE HISTORY NO. 6 - PLANT DESIGN CONDITIONS

Description	Original Plant 1953	1957	1964	1979	Current Upgrading
Design Average Flow, mgd	2.25	8.0	8.0	8.6	10.7
Preaeration Tanks-					
Detention Time (DAF[1]), minutes	–	–	–	–	14.8
Primary Settling Tanks-					
Detention Time (DAF), hr	2.7	2.1	2.1	2.1	1.6
Overflow Rate (DAF), gpd/sq ft	670	1210	1210	820	1,000
Secondary Treatment-					
Primary Trickling Filters-					
Organic Loading (DAF), lb BOD/day/1,000 cu ft	12	17			–
Hydraulic Loading (DAF), mgd/acre[2]	1.8	2.2			–
Second Trickling Filters-					
Organic Loading (DAF), lb BOD/day/1,000 cu ft	–	33	62	62	85
Hydraulic Loading (DAF), mgd/acre[2]	–	3.6	9.2	9.2	12.4
Aeration Tanks-					
Detention Time (DAF), hr	–	–	7.3	14.7	11.8
Final Clarifiers-					
Detention Time (DAF), hr	1.4	3.5	3.5	3.5	2.8
Overflow Rate (DAF), gpd/sq ft	1,200	460	460	460	620
Chlorination Tanks-					
Contact Time (DAF), minutes	–	38	38	38	5

TABLE 11-9 (Continued)

Description	Original Plant 1932	1957	1964	1976	Current Upgrading
Sludge Thickening					
Flotation Unit Surface Area, sq ft					250
Sludge Digestion (Number of Units)					
60 ft diameter x 22 ft SWD anaerobic	1	1	1		
80 ft diameter x 20 ft SWD anaerobic		1	1		
Sludge Dewatering					
Sludge Drying Beds, sq ft	22,600				
Vacuum Filter Surface Area, sq ft		230	230	230	230
Ultimate Disposal	Landfill	Landfill	Landfill	Incineration	Incineration

[1] DAF - Design Average Flow.

[2] Does not include recirculation.

TABLE [illegible]

SUMMARY OF THE [illegible] PERFORMANCE [illegible]

	[illegible]		[illegible]		[illegible]		
	Before Upgrading	After Upgrading	Before Upgrading	After Upgrading	Before Upgrading	After Upgrading	Current Upgrading
Average Flow, mgd	[illegible]	[illegible]	[illegible]	[illegible]	[illegible]	[illegible]	[illegible]
Peak Flow, mgd	[illegible]	[illegible]	[illegible]	[illegible]	[illegible]	[illegible]	[illegible]
BOD							
Raw Wastewater, mg/l	[illegible]	[illegible]	[illegible]	[illegible]	[illegible]	[illegible]	[illegible]
Removal, percent	[illegible]	[illegible]	[illegible]	[illegible]	[illegible]	[illegible]	[illegible]
SS							
Raw Wastewater, mg/l	[illegible]	[illegible]	[illegible]	[illegible]	[illegible]	[illegible]	[illegible]
Removal, percent	[illegible]	[illegible]	[illegible]	[illegible]	[illegible]	[illegible]	[illegible]

TABLE 13-11
CAPITAL COSTS OF UPGRADING FOR CASE HISTORY NO. 6

Date of Upgrading	Item	Capital Cost Cost at Time of Upgrading	Capital Cost 1973 Cost
1957	Screening and Degritting	$ 45,000	$ 110,000
	Rehabilitation of Power and Control Building	150,000	365,000
	Primary Clarification	65,000	160,000
	Secondary Process	400,000	1,000,000
	Final Clarification	115,000	280,000
	Chlorination Facilities	25,000	60,000
	Sludge Handling	300,000	730,000
	Miscellaneous	150,000	365,000
	Total	$1,250,000	$2,930,000
1964	Modifications to Secondary Process	200,000	367,000
1970	Modifications to Secondary Process	100,000	133,000
Current Design	Renovation of Screens and Degritting Equipment		60,000
	Preaeration and Odor Control		160,000
	Enclosure of Primary Tanks and Odor Control		200,000
	Phosphorus Removal Facilities		75,000
	Effluent Filters and Aerators		450,000
	Sludge Handling		250,000
	Miscellaneous		55,000
	Total		$1,250,000

13.7 Case History No. 7 - Upgrading by Optimization of Aeration Tank-Clarifier Relationship (9)

The Coldwater Creek Wastewater Treatment Plant in St. Louis, Missouri is an activated sludge plant designed to treat a flow of 25 mgd. The plant facilities include six aeration tanks and four final clarifiers; however, at the time of upgrading, only three of the aeration tanks were being used to treat a flow of 14.2 mgd. Under these operating conditions, the plant efficiency was poor with BOD and SS removals of 73 and 46 percent, respectively.

Table 13-13 summarizes operating data before and after optimization of the aeration tank-clarifier relationship. The flow diagrams presented on Figure 13-7 depict the changes in operation that were made. Analysis of the data indicates that before the modifications, the return sludge rate and the resulting MLSS concentration in the aeration tanks were both high. These conditions imposed a solids loading rate on the final clarifiers in excess of the recommended range (20-30 lb/day/sq ft) shown in Table 6-2, even though the overflow rate was within acceptable limits. As a result, poor clarifier performance lowered the overall BOD and SS removals normally expected in an activated sludge system.

TABLE 13-13
CASE HISTORY NO. 7 - OPERATING DATA

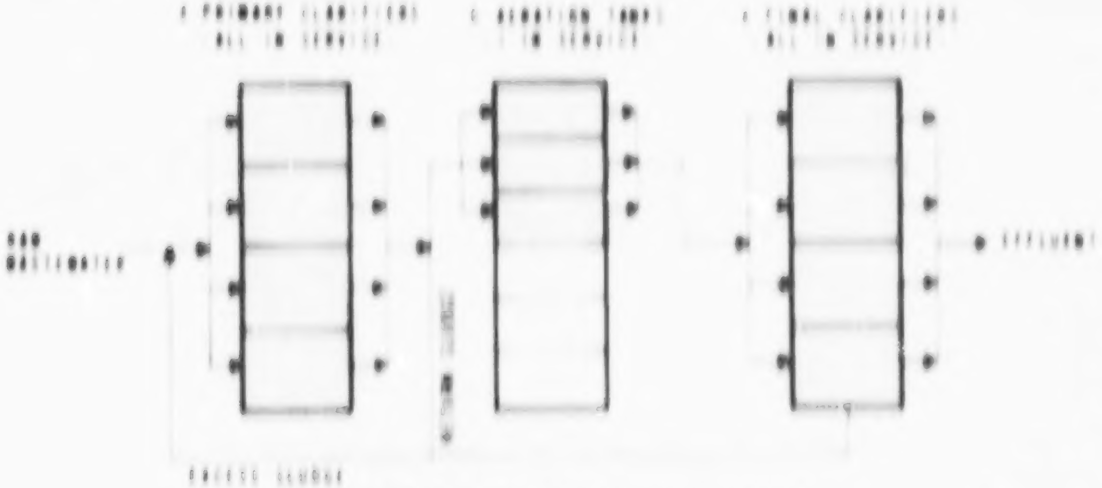
Description	Before Operating Modifications	After Operating Modifications
Average Flow, mgd	13.2	13.5
Aeration Tanks:		
Tanks in Service	3	4
MLSS, mg/l	7,200	3,400
Return Sludge Rate, percent of Average Flow	93	40
Volumetric Loading, lb BOD/day/1,000 cu ft	43.7	33.5
Organic Loading, lb BOD/day/lb MLSS	0.10	0.15
Final Clarifiers:		
Tanks in Service	4	3
Overflow Rate at Average Flow, gpd/sq ft	515	755
Solids Loading Rate at Average Flow	59.0	27.0

To improve plant performance, a fourth aeration tank was put in service. A lower return sludge rate was used and the MLSS concentration in the aeration tanks was decreased by over 50 percent. With these operational modifications, it was possible to maintain both the final clarifier solids loading rate and the overflow rate within their recommended limits with only three final clarifiers in service. The improvement in treatment plant efficiency resulting from the proper balancing of the aeration tanks and final clarifiers is shown in Table 13-14.

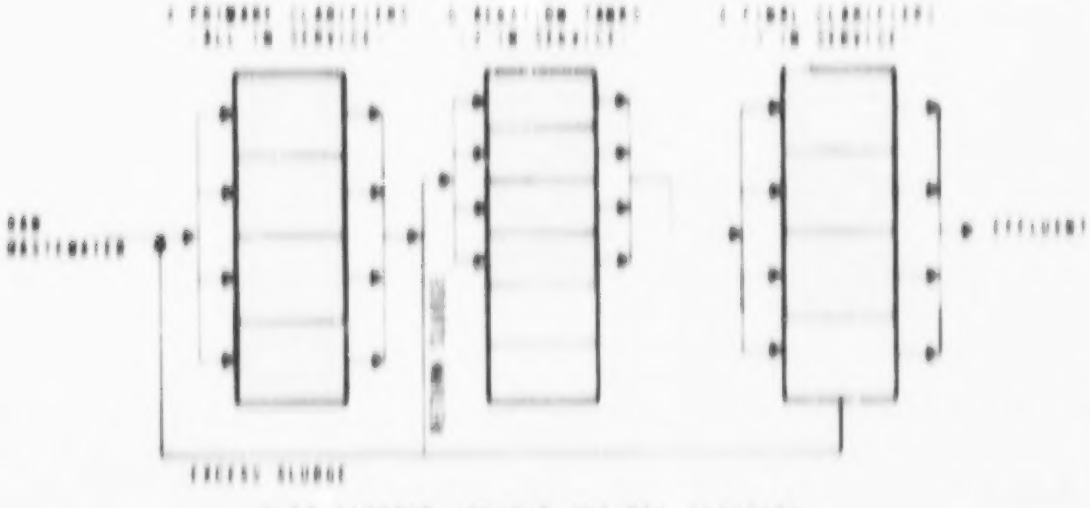
FIGURE 13.7

CASE HISTORY NO 7

COMPARISON OF ORIGINAL AND MODIFIED FLOW DIAGRAMS


[illegible] PRIMARY CLARIFIERS
ALL IN SERVICE
[illegible] AERATION TANKS
[illegible] IN SERVICE
[illegible] FINAL CLARIFIERS
ALL IN SERVICE
RAW WASTEWATER
EFFLUENT
RETURN SLUDGE
EXCESS SLUDGE

FLOW DIAGRAM BEFORE OPERATING MODIFICATION


[illegible] PRIMARY CLARIFIERS
ALL IN SERVICE
[illegible] AERATION TANKS
[illegible] IN SERVICE
[illegible] FINAL CLARIFIERS
[illegible] IN SERVICE
RAW WASTEWATER
EFFLUENT
RETURN SLUDGE
EXCESS SLUDGE

FLOW DIAGRAM AFTER PLANT MODIFICATION

TABLE 14-13
CASE HISTORY NO. 7 - PERFORMANCE DATA

Description	Before Modification of Flow Pattern	After Modification of Flow Pattern
Average Wastewater Flow, mgd	4.2	4.5
BOD		
Raw Wastewater, mg/l	150	162
Primary Effluent, mg/l	132	140
Final Effluent, mg/l	38	9†
Overall Plant Removal, percent	75	94
SS		
Raw Wastewater, mg/l	173	190
Primary Effluent, mg/l	135	113
Final Effluent, mg/l	92	16†
Overall Plant Removal, percent	47	92

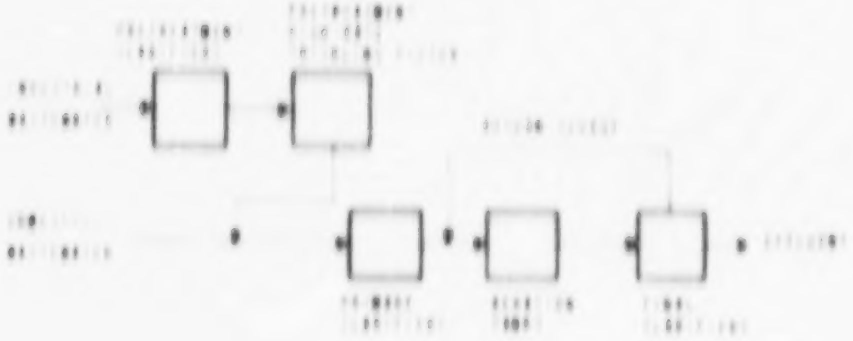
† One week average.

14-8 Case History No. 8 - Upgrading by Optimization of Aeration Tank-Clarifier Relationship (9)

The Sioux Falls, South Dakota Wastewater Treatment Plant is a 9.5 mgd activated sludge plant. Approximately 4.5 mgd of the flow is industrial wastes (meat packing wastes with a BOD of 2,500 mg/l). The industrial wastes are pretreated by clarification and high-rate trickling filtration. The pretreated industrial wastes are then added to the main wastewater flow, which is primarily domestic, and the combined flow is treated by primary clarification followed by an activated sludge process. A flow diagram of the major treatment processes is presented on Figure 14-8.

During most of the year, the plant provided excellent treatment. However, during the fall, the treatment plant is subjected to high industrial loads that significantly lowered plant efficiency prior to optimization of the aeration tank-clarifier relationship. Table 14-14 shows operating and performance data for the months of October and November, the period when loads are highest. The aeration tank loadings during 1967 were extremely high, while the solids loading rate on the final clarifiers was low. The unbalanced loadings on the aeration tanks and final clarifiers were attributed to the relatively low return sludge rate and the resulting low MLSS concentration.

FIGURE 13-8
CASE HISTORY NO. 8
FLOW DIAGRAM



To improve treatment plant efficiency during the peak load season, it was decided to lower the organic loading on the aeration tanks by increasing the return sludge rate, thereby raising the MLSS concentration. The success of this operational modification was dependent on the ability of the final clarifiers to satisfactorily handle the increased solids loading. Table 13-14 also presents operating and performance data for October and November, 1968, after implementation of the modified operational mode. These data indicate that the final clarifier solids loading rate was substantially raised, although still at an acceptable level, without affecting the overflow rate. The improvements to treatment plant efficiency are evident in Table 13-14.

TABLE 13-11
CASE HISTORY NO. 8 OPERATING AND PERFORMANCE DATA

Description	1967 Before Modification of Operating Mode	1968 After Modification of Operating Mode
Average Wastewater Flow, mgd	10.0	9.8
Aeration Tank:		
MLSS, mg/l	1,800	1,800
Return Sludge Rate, percent of Average Flow	30	90
Volumetric Loading, lb BOD/day/1,000 cu ft	115	117
Organic Loading, lb BOD/day/lb MLSS	1.04	0.63
Final Clarifier:		
Overflow Rate, gpd/sq ft	600	650
Solids Loading Rate, lb/day/sq ft	7	11
Final Effluent[1]		
BOD, mg/l	29.5	20.0
SS, mg/l	14.5	13.5

[1] Fifty percent of the time effluent quality was equal to or less than value shown.

13.9 Case History No. 9 Upgrading a Modified Aeration System for Nutrient Removal (10)

In 1969, regulatory agencies established more stringent effluent standards for treatment plants that discharge into the Potomac River in the vicinity of Washington, D.C. These higher standards required upgrading the Washington, D.C. Blue Plains Plant to provide phosphorus and nitrogen removal as well as improved BOD and SS removals.

In 1969, very little performance data were available on the alternative phosphorus and nitrogen removal methods that might be used in this upgrading situation. Fortunately, through the cooperation of the Joint EPA-DC Pilot Plant, it was possible to pilot and evaluate several alternative nutrient removal treatment sequences. Based on these studies, two-point addition of a metal salt was selected for phosphorus removal, and it was determined that nitrogen removal would be best achieved through biological nitrification and denitrification processes. The pilot studies also indicated that to consistently meet the established effluent standards, multimedia filtration would be required. Anticipated performance data for the upgraded plant are presented in Table 13-15. Figures 13-9, 13-10, and 13-11 show, respectively, the flow diagrams for the primary and secondary systems, nitrification and denitrification systems, and filtration and disinfection systems of the upgraded plant.

The existing secondary system consists of four aeration tanks and 12 sedimentation units. To handle the anticipated increase in plant design flow from 240 mgd to 309 mgd, the existing secondary system will be upgraded with two additional aeration tanks and 12 additional final sedimentation tanks. The aeration tanks are designed for a volumetric loading of 120 lb BOD/day/1,000 cu ft, an organic loading of 2 lb BOD/day/lb MLSS, and a MLSS concentration of 1,000 mg/l (10). Since the increased design loadings require more air per unit volume than the existing aeration system can deliver, the existing aeration system will be expanded. This system will be modified from a coarse bubble, spiral roll system to a coarse bubble, grid pattern to improve oxygen transfer efficiency. The secondary system air capacity has been designed to provide 0.51 lb O_2/lb BOD removed (10).

Alum or ferric chloride will be added to the mixed liquor of the secondary system and is expected to remove approximately 70 percent of the phosphorus contained in the plant influent. To remove most of the remaining phosphorus, metallic salts will also be added to the nitrogen release tanks. The addition of metallic salts to the secondary system is also expected to improve the BOD removal in the secondary system from 75 percent to 85 percent. This will ensure a secondary effluent BOD concentration of less than 40 mg/l, which was found during pilot testing to be desirable for successful nitrification in the second stage.

Biological nitrification facilities are designed for oxidation of 0.066 lb NH_3-N/day/lb MLVSS at minimum wastewater temperatures and a MLVSS concentration of 1,700 mg/l. At the stoichiometric oxygen requirement of 4.6 lb O_2/lb NH_3-N oxidized, 120 75-hp turbine aerators are required (10). Maximum air supply to the turbines will be 100,000 cfm. The turbines were selected in this instance because, due to the limitations of the site, the nitrification tanks are designed to have depths of 30 feet to obtain the required volume. The turbines will provide adequate mixing to this depth and are capable of supplying a wide range of oxygen to the system as required by varying NH_3-N influent loads and varying wastewater temperatures. Lime will be added to the nitrification reactor to maintain the necessary pH for nitrification. The nitrification sedimentation tanks are designed for average and peak hydraulic and solids loadings of 500 and 1,210 gpd/sq ft and 17.4 and 36.6 lb/sq ft/day, respectively. The sludge return system is designed to provide return of

TABLE 11-17

CASE HISTORY NO. 9 - ANTICIPATED PERFORMANCE AFTER UPGRADING

	Grit Chamber Effluent	Secondary Effluent[1]	Nitrification Effluent	Denitrification Effluent	Filtration Effluent	Effluent Standard
BOD, mg/l	200	25	10	6	2	5
Total Phosphorus, mg/l	8.3	2.0	1.0	0.5	0.2	0.25
Nitrogen:						
Organic-N, mg/l	8.6	2.0	1.0	1.0	0.5	
NH_3-N, mg/l	13.7	12.8	1.5	1.0	1.0	
NO_2+NO_3-N, mg/l	0	0.2	11.1	1.0	0.5	
Total N, mg/l	22.3	15.0	13.6	3.0	2.0	2.0

[1] With ferric chloride or alum addition.

FIGURE 13.8

UPGRADING A MODIFIED AERATION SYSTEM FOR NUTRIENT REMOVAL

FLOW DIAGRAM

PRIMARY AND SECONDARY SYSTEMS (1)

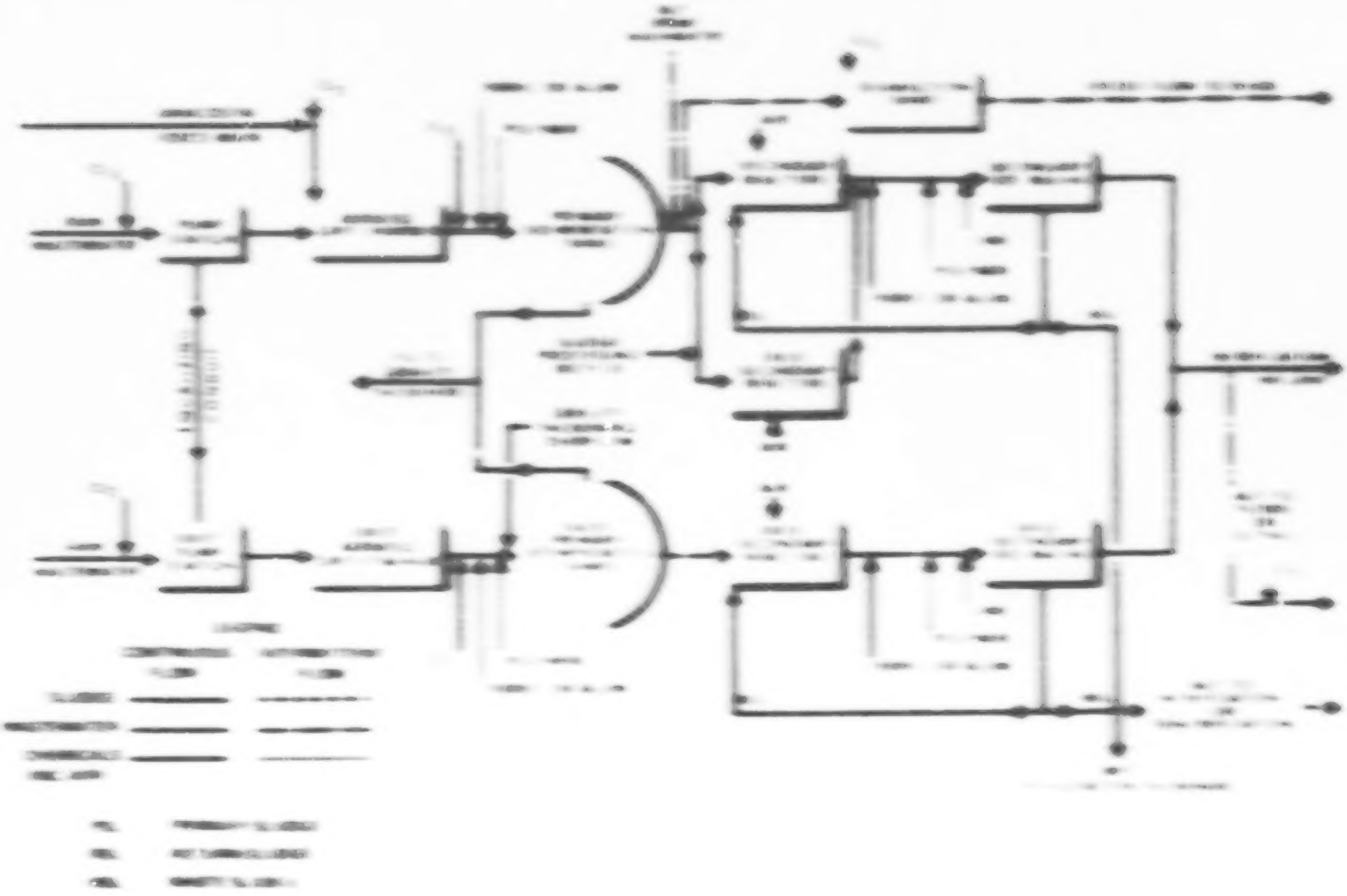

LEGEND
CONTINUOUS FLOW
INTERMITTENT FLOW
SLUDGE
WASTEWATER
CHEMICALS
AND AIR
PS PRIMARY SLUDGE
RS RETURN SLUDGE
WS WASTE SLUDGE

FIGURE 13-10

UPGRADING A MODIFIED AERATION SYSTEM FOR NUTRIENT REMOVAL

FLOW DIAGRAM

NITRIFICATION AND DENITRIFICATION SYSTEMS (10)

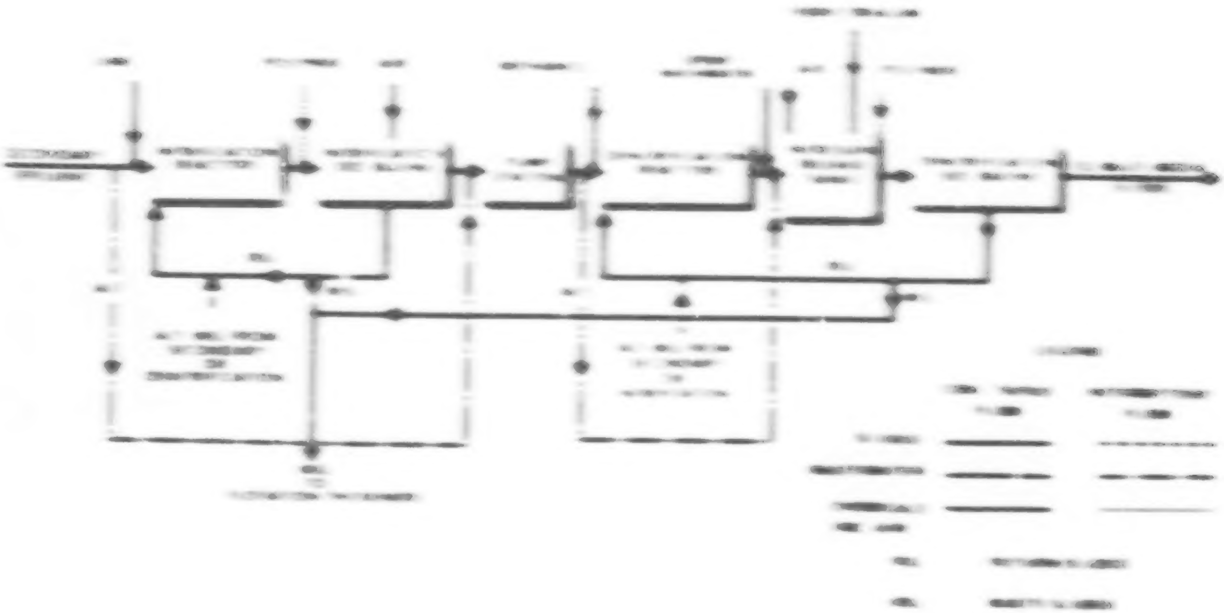
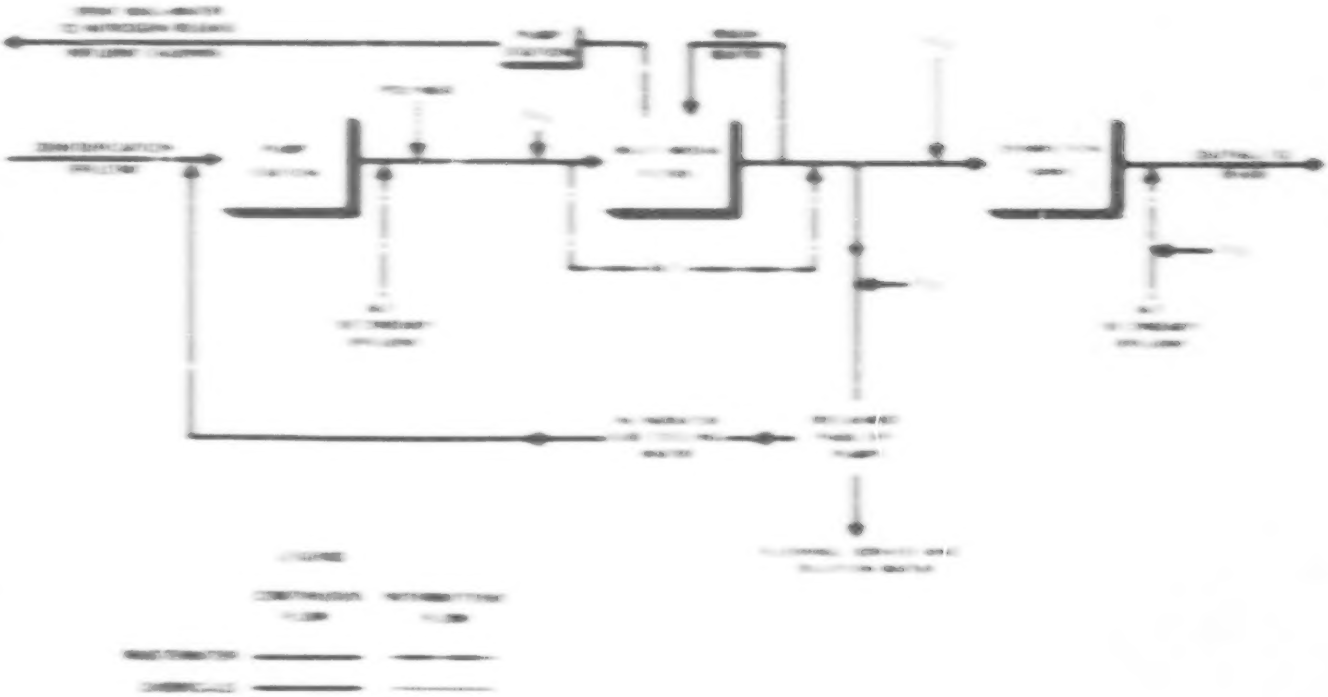


FIGURE 13.11

UPGRADING A MODIFIED AERATION SYSTEM FOR NUTRIENT REMOVAL
FLOW DIAGRAM
FILTRATION AND DISINFECTION SYSTEMS (3)



DENITRIFICATION EFFLUENT
PUMP STATION
POLYMER
OUTFALL TO RIVER
LEGEND
CONTINUOUS FLOW
INTERMITTENT FLOW
WASTEWATER
CHEMICALS

100 percent of peak flow. However, the system will normally be operated to return 100 percent of the average flow. Continuous monitoring of the DO content of the nitrification effluent will be provided to ensure that the influent DO to the denitrification system is minimized.

The biological denitrification system will include reactors, nitrogen release tanks and sedimentation tanks. The reactors have been designed for removal of 0.0125 lb NO_3-N/day/lb MLVSS at a design MLVSS concentration of 2,100 mg/l, with up to 4.5 lb methanol added/lb NO_3-N applied (10). The reactors will be equipped with 10.5 hp mixers and will be covered but not airtight.

The nitrogen release tanks will serve three functions: (1) to strip supersaturated nitrogen gas, (2) to provide mixing for second-stage metal salt addition for residual phosphorus removal, and (3) to provide an aeration zone for removal of excess methanol. These tanks will furnish a 23-minute detention period at average flow.

The denitrification sedimentation tanks are designed for hydraulic loadings of 670 gpd/sq ft at average flow and 1,440 gpd/sq ft at peak flow. Solids loadings are 25.6 and 54.0 lb/sq ft/day at average and peak flows, respectively.

The 46 multimedia filters are designed for filtration rates of 3.0 gpm/sq ft at average flow and 6.2 gpm/sq ft at peak flow. Backwashing will occur at normal intervals of 24 hours at a rate of 25 gpm/sq ft. The backwash water will be equalized in conduits and may be returned upstream of either the secondary reactors or the nitrogen release tanks.

Provision has been made to chlorinate either upstream or downstream of the filters with 23 minutes detention provided in contact tanks following the filters.

Sludge processing facilities will include gravity thickening of primary sludge, flotation thickening of secondary and advanced treatment sludges, vacuum filtration and sludge incineration.

13.10 References

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3. Private communication with J.D. Wittenmeyer, Vice President, Ohio Sanitation Service Company, Dayton, Ohio (January 22, 1971).

4. Burns & McDonnell Engineering Company. Report on Sewage Treatment Plant and Sanitary Sewer Improvements for Sedalia, Missouri (1963).

5. Wirts, J.J., *The Use of Organic Polyelectrolyte for Operational Improvement of Waste Treatment Processes.* Federal Water Pollution Control Administration, Grant No. WPRD 102-01-68 (May, 1969).

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9. West, A.W., *Case Histories of Plant Improvement by Operations Control, Nutrient Removal and Advanced Waste Treatment.* Federal Water Pollution Control Administration, Cincinnati, Ohio (1969).

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APPENDIX A
METRIC CONVERSION CHART

Multiply	By	To Get
Inches	2.54	Centimeters
Feet	0.3048	Meters
Square Feet	0.0929	Square Meters
Cubic Feet	0.0283	Cubic Meters
Pounds	0.454	Kilograms
Gallons	3.79	Liters
Gallons/Minute	5.450	Cubic Meters/Day
Feet/Second	0.305	Meters/Second

APPENDIX B

WORD ABBREVIATIONS

	Text	Tables
afternoon	PM	PM
average	--	avg
before noon	AM	AM
biochemical oxygen demand	BOD	BOD
brake horsepower	BHP	BHP
centimeter(s)	cm	cm
chemical oxygen demand	COD	COD
cubic foot	cu ft	cu ft
cubic feet per minute	cfm	cfm
cubic feet per second	cfs	cfs
degree Centigrade	deg C	°C
degree Fahrenheit	deg F	°F
dissolved oxygen	DO	DO
elevation	El	El
feet per minute	fpm	fpm
food to microorganism ratio	F/M	F/M
foot (feet)	--	ft
gallon(s)	--	gal
gallons per day	gpd	gpd
gallons per minute	gpm	gpm
gram(s)	--	g
head loss	--	H L
horsepower	hp	hp
hour(s)	--	hr
inch(es)	--	in
Jackson turbidity units	--	jtu
kilowatt hour	kwh	kwh
micron(s)	μ	μ
milligrams per liter	mg/l	mg/l
milliliter	ml	ml
millimeter(s)	mm	mm
million gallons	mil gal	mil gal
million gallons per day	--	mgd

APPENDIX B – Continued

	Text	Tables
mixed liquor suspended solids	MLSS	MLSS
mixed liquor volatile suspended solids	MLVSS	MLVSS
most probable number	--	MPN
number(s)	--	No.
parts per million	ppm	ppm
pound(s) (weight)	--	lb
pounds per cubic foot	pcf	pcf
pounds per square foot	psf	psf
pounds per square inch gage	psig	psig
revolutions per minute	rpm	rpm
second(s)	--	sec
side water depth	--	SWD
sludge retention time	SRT	SRT
square foot	sq ft	sq ft
standard oxygen transfer rate	SOR	SOR
suspended solids	SS	SS
total solids	--	TS
total suspended solids	--	TSS
volatile suspended solids	--	VSS
year(s)	--	yr